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STUDY OF THE CHARACTERIZATION AND CONTROL OF AIR POLLUTANTS FROM A FLUIDIZED-BED BOILER--THE SO2 ACCEPTOR PROCESS

J. S. Gordon, et al

Pope, Evans and Robbins, Incorporated

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STUDY OF THE CHARACTERIZATION AND CONTROL OF AIR POLLUTANTS FROM A FLUIDIZED-BED BOILER–THE SO_2 ACCEPTOR PROCESS

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15. Supplementary Notes

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16. Abstracts The report describes the development of the SO2 Acceptor Process, an atmospheric pressure, coal-fired fluidized-bed boiler concept for steam and power generation. Coal is burned in a fluidized bed of crushed lime in the boiler's primary combustion zone; the partially sulfated lime is continuously regenerated (by reductive decomposition) in another zone. High combustion efficiencies are achieved by recycling, to a Carbon Burnup Cell, the carbon-containing flyash which is carried out of the primary combustion zone. Experimental work was conducted in a 100 lb coal/hr batch combustor, and in a continuous 800 lb coal/hr pilot boiler, for the purpose of demonstrating system operability, including: high degrees of sulfur removal in the primary combustion zone; high levels of SO2 in the off-gases from the regeneration zone, suitable for sulfur recovery; and high combustion efficiencies. Preliminary designs and cost estimates are presented for 30-MW and 300-MW boilers.

17. Key Words and Document Analysis. 17a. Descriptors

Air Pollution
Fluidized Bed Processing
Sulfur Oxides
Nitrogen Oxides
Limestone
Combustion
Coal
Calcium Sulfates
Sulfur
Additives
Fossil Fuels
Calcium Oxides
Stoichiometry
Regeneration (Engineering)

17b. Identifiers/Open-Ended Terms

Air Pollution Control
Stationary Sources
SO2 Acceptor Process
Calcium Sulfate Regeneration
Fluidized-Bed Combustion

17c. COSATI Field/Group 13B, 07A

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1. SUMMARY

1.1 General

Work is described on studies leading to low-cost environmentally-clean power generation from coal. A regenerative limestone process for multi-cell fluidized-bed coal combustion-desulfurization is described.

Pope, Evans and Robbins (PER), in the final phase of a project* sponsored by the Office of Research and Monitoring, Environmental Protection Agency, has studied the air pollutant emissions reduction capability of fluidized-bed boilers for the combustion of coal. In earlier work with sintered ash beds, it was found that sulfur-oxide emissions could be markedly reduced by injecting finely divided limestone on a once-through basis into a coal-burning, atmospheric fluidized-bed operating at 1500°F to 1600°F#, and with about 3% residual oxygen or more remaining in the flue gas.** At 3% residual oxygen or less in the flue gas, a significant percentage of the input coal's fuel value would appear as carbon in the boiler fly ash. A comprehensive search for methods to reduce the elutriation loss of fuel, which ranged to 15% for the high air rates (superficial velocity about 9 ft/sec) used in the PER designs, led to the invention of the Carbon-Burnup Cell.*** The alternative, elutriation minimized by operation at low gas velocity, was shown to be uneconomical due to increased boiler size. The Carbon-Burnup Cell was simply a region of a fluidized-bed boiler into which carryover from the primary boiler bed was injected, and in which the amount of heat transfer surface per unit heat release was lower than in the adjacent coal-fired regions. In this way a temperature in the 1900°F to 2050°F range could be maintained. Design criteria for an effective Carbon-Burnup Cell were produced in the previous phase of this program*. Reinjection of carbon-bearing fly ash into the primary boiler bed was shown inferior to the Carbon-Burnup Cell technique, since the fly ash carbon must compete with fresh coal for the available oxygen.

---

* This report describes the results of experiments carried out between November 1970 and August 1971. Primary emphasis has been placed on bituminous coal combustion.

** Reference 2 describes the details of the sulfur control program.

*** U.S. Patent 3,508,506.

# Although it is EPA policy to use metric units, certain non-metric units are used in this report to reflect actual test results. Please use the conversion factors in Appendix B if you are more familiar with the metric system.

It was desired to improve upon the sulfur capture results obtained with once-through limestone injection into a coal-burning, sintered ash fluidized bed. Tests PER ran in May 1969 showed that limestone was a suitable bed material in place of the sintered inert ash. Lime bed operation was found promising for SO_2 capture. A continuous capture/regeneration operation of the sorbent was visualized. PER also found that injection of salt (sodium chloride) aided lime SO_2 capture effectiveness and also increased combustion efficiency. The major goal of the study described in this report was to investigate these types of process operations.

Therefore, the program described in this report consisted of two major tasks: laboratory scale, batch type coal combustion experiments using limestone beds for sulfur capture, with and without salt additive to modify lime effectiveness; and pilot plant (boiler system) experiments in which the limestone from the primary combustion zone is made to undergo continuous regeneration in a regeneration section. The primary benefit of the regenerator technique is to minimize makeup limestone requirements at continuous high sulfur capture levels in the boiler. Without regeneration, large quantities of limestone would have to be cross-flowed through the boiler and discarded. A second advantage of this regeneration technique is that the regeneration zone produces a SO_2-rich flue gas stream, which is volumetrically only a small fraction of the total boiler gas flow; this SO_2-rich gas may be fed to a sulfur recovery operation.

---

* Reference 1.

The regeneration scheme of the SO_2 Acceptor Process works as follows: sulfated lime from the 1500°F primary zone continuously circulates to the 2000°F regeneration zone where carbon-bearing fly ash (MK I) or coal (MK II) is burned with low excess air. As described in Refs. 1 and 2, SO_2 is driven off in high concentration and lime is recovered for reuse by the reaction:

$$CaSO_4 + C + \tfrac{1}{2}O_2 \rightarrow CaO + SO_2 + CO_2$$

In the SO_2 Acceptor Process MK I configuration, the regeneration is performed in the Carbon-Burnup Cell and the process consists of two zones. In the MK II configuration, separate regenerator and Carbon-Burnup Cells are used, and the process consists of three zones.

The goals of the present pilot plant study were:

a. 90% or better SO_2 removal from primary cell
b. 3-4% or more SO_2 in regenerator flue gas
c. 98% or better carbon burnup in the system

In these tests, the regenerator was fly ash or coal-fired. Use of fuel gas for $CaSO_4$ regeneration has been avoided for economic reasons. Because of current natural gas shortages, there is little possibility that a gas-fired regenerator could be guaranteed a fuel supply.

Use of pulverized coal fuel has been avoided in order to achieve high combustion efficiency, as well as economy of coal preparation. Superficial gas velocities have been above 6 ft/sec for economic reasons, i.e., to minimize boiler size.

1.2 Tests in the Fluidized-Bed Column

The purpose of the FBC tests was to explore SO_2 capture and salt addition on a bench scale prior to a major expenditure of resources in FBM testing. The first portion of the effort, to explore limestone bed effectiveness for SO_2 capture, was carried out in the Fluidized-Bed Column (FBC) which, during these tests, had a plan area of 0.86 square feet. FBC operation in this program was nonregenerative. In previous programs, the FBC had been used in cyclic sorption-regeneration tests.

Coal was burned in this device over a wide range of conditions, summarized as follows:

Bed temperature:	1470°F to 1650°F
Bed depth (static):	14" to 22"
Air rate:	700 to 800 lb/hr ft^2
Fuel rate:	47 to 81 lb/hr ft^2

The method of FBC operation was: the initial bed for each test was fresh limestone at a low bed level (about 8 inches). The bed was then brought to calcining temperature by the gas burner and coal feed. More bed was then added to create the desired depth. A calcining period of several hours thus precedes the start of SO_2 release buildup. If the gas burner light-off had been attempted with the full bed depth, preparation time would have been much longer.

Flue gas analyses, sulfur balances, and bed particle size vs. time analyses were compiled for these tests. Portions of the operation utilized salt addition to provide lime activity enhancement as well as combustion catalysis in a nonregenerative operation.

Typical results of FBC tests are: carbon burnup in one pass is typically 90 to 96%; bed sulfur content increases with time since the operation is nonregenerative; as the bed deactivates, the gas SO_2 level also increases from 200 ppm to 1000 or more in the absence of salt; SO_2 levels with the aged lime bed are reduced dramatically by salt addition. The fate of the salt is to produce more fly ash, since NaCl is believed to be converted to $Na_2Ca(SO_4)_2$ and the gravimetric factor, $CaNa_2(SO_4)_2/NaCl$, is 4.7. The chlorine fed is believed to be tied up by lime as $CaCl_2$. No equipment corrosion due to salt has been observed. The vapor pressure of NaCl as a function of temperature is listed in various sources; e.g., the JANAF Thermochemical Tables. At 1500°F, it is significant and NaCl acts as a homogenous vapor phase combustion catalyst. Also, the isokinetic samples in the cooled flue gas stream display a high salt fog content. Solid salt feeder operation was unreliable. Good results were obtained with aqueous salt solution injection, but the water vaporization thermal penalty is undesirable.

1.3 Tests in the Pilot Plant (Fluidized-Bed Module)

Several tests of longer duration were conducted in the Fluidized-Bed Module (FBM), a boiler with an air distributor grid area of 9 ft^2. Its rating is 5000 lb/hr steam, 800 lb/hr coal input. Two different designs for Carbon-Burnup Cells were appended to the FBM to determine the problem areas in operating a regeneration-cycle fluidized bed with two distinct temperature zones (1500°F in FBM, 2000°F in CBC). A vertical coal feeder, with flow splitter and two horizontal opposed outlets, was used in the FBM. Some operating problems occurred with this feeder. Coal feeding to a fluidized-bed boiler is an area which requires further development effort.

As discussed in Reference 1, an experimental Carbon-Burnup Cell was appended to the FBM, to bring system carbon burnup into the 98-100% range. Without the CBC, the FBM provides 90% carbon burnup in one pass. CBC emissions of hydrocarbons and carbon monoxide were essentially zero when operating in the Carbon-Burnup Cell mode. It was found that for the small CBC ($\sim$1.1 ft^2) an opening in the barrier between the FBM and CBC regions of just 2 square inches permitted the desired temperature difference to be achieved. An interchange rate between the FBM and CBC of 12,000 pounds of bed material per hour per square foot of opening was estimated from two transient heating tests described in Reference 1. When the CBC was used as a regenerator, interchange was sometimes supplemented by screw feeding bed material from the FBM to the CBC, by which means the CBC flue gas SO_2 content was increased.

Two CBC configurations were used in the current program, both incorporating bed particle "knockouts" or "baffle screens". Figures 7 through 10 show most of the details of the two CBC's. The baffle screens consisted of cylinders arranged in a triangular array. With the short CBC configuration*, without a baffle screen, large quantities of bed material had been expelled from the CBC. With either baffle screen, the quantity of bed material carried over was small. With the short CBC (Figure 9) configuration, when the baffle screen was made up of water-cooled tubes, an initial* heat transfer coefficient of ~35 Btu/ft^2 hr°F was measured. This was approximately twice the value that would be predicted from simple radiation and convection, indicating that an active heat transfer region exists above the dense phase of a high velocity (hence turbulent) fluidized bed. With the tall CBC shown in Figure 10, the coefficient was only 20 Btu/ft^2 hr°F where the temperature is the log mean temperature differential**. The lower coefficient is caused by the fact that the baffle screen in the tall CBC was far above the fluid bed.

---

* This original short configuration resulted from the FBM steam drum layout. It is shown in Figures 7 and 9.

When all operational problems has been overcome, the FBM with the tall CBC achieved an overall carbon combustion efficiency of 99%.

When the CBC was used as a regenerator, the rate of regeneration of bed material was shown to be increased by coal feed to the CBC; but this procedure reduced carbon combustion efficiency of the system below 98%.

The integrated FBM/CBC boiler system was operated extensively to investigate process parameters of the two-cell "SO_2 Acceptor Process," labelled the MK I system. First, a shakedown test of about 28 hours' duration was made. An extended run of about 80 hours' duration was then made. In the MK I system, coal is burned in the FBM where an active lime bed (-8+20 mesh) at the optimum sorption temperature (1550°F) absorbs better than 90% of the SO_2 emitted by coal combustion. Carbon-bearing fly ash from the FBM is collected and introduced to the high-temperature CBC where further carbon burnup occurs. As a function of CBC fuel and air rates, temperature and combustion efficiency, FBM bed material circulating through the CBC is regenerated. Three benefits result:

---

* Prior to onset of noticeable fouling buildup.

** Temperatures and rates of flue gas and cooling water are measured entering and leaving the cooler.

(1) A relatively SO_2-rich flue gas stream is synthesized in the CBC which volumetrically is only a small fraction of the total system flue gas.

(2) Lime containing $CaSO_4$ is regenerated, minimizing $CaCO_3$ makeup and calcining requirements in the FBM.

(3) FBM bed density is minimized*, aiding efficient distribution of coal within the aerated bed.

Coal containing 4.5% sulfur was used, limiting the CBC gas SO_2 content to a maximum of about 3%**. Additional process parametric variations were studied during these tests: coal addition to the CBC/Regenerator to supplement reducing conditions provided by fly ash feed; salt addition to the FBM to enhance SO_2 capture and combustion efficiency.

Following these tests, two additional tests of about 8 hours' duration each were made in an effort to reduce CBC air rate and hence increase SO_2 concentration. CBC operation was on coal only, to aid regeneration at very low CBC air rates. Fly ash feed was restricted to the CBC to minimize CBC air rate requirements; salt was added to the FBM to enhance carbon burnup, thereby reducing the CBC carbon burnup requirement and associated air rate.

---

* A bed of calcium sulfate is denser than one of calcium oxide.

** The SO_2 concentration in the CBC off gas is set by the mass balance and could come close to the thermodynamic equilibrium concentration of 8 to 11%.

The conclusions regarding two-cell operation of the SO_2 Acceptor Process, MK I, drawn from these tests are:

(1) The tandem FBM/CBC system operates in the SO_2 Acceptor Process mode continuously and with stability.

(2) The SO_2 removal from the primary cell (FBM) flue gas, as determined by continuous infrared analysis, is easily maintained in the 90-95% range.

(3) Limestone makeup requirements are remarkably low, about 5% of coal input by weight. The same bed was used in a month of testing, with additions needed to make up for analytical samples withdrawn and attrition losses which are very modest with the 1359 limestone. No limit on the sorption-regeneration cycles tolerated by the lime bed has been observed, and activity remains at a high level.

(4) The coal combustion efficiency of the overall system can be maintained for an extended period at 98% or better using FBM/CBC burnup conditions comparable to testing earlier in this Contract*. Under these conditions, a minimum amount of, or preferably no, coal is fed to the CBC, O_2 levels greater then 1% are maintained in the CBC off-gases, and SO_2 levels in the CBC gas are below 3%, using the 4.5% S coal. Under these conditions, sulfur balances

---

* Reference 1

(by analysis of gas streams, ash, and bed material) have not been of high accuracy, and incomplete regeneration of bed material is probably occurring due to the need for O_2 levels of more than 1%. the CBC gas as imposed by burnup demands.

(5) The SO_2 concentration in the CBC off-gases, also as determined continuously by infrared, can be maintained at levels of 3-4%, corresponding to tandem operation in the SO_2 Acceptor Process MK I mode, for an extended period using less oxidizing conditions in the CBC (0.5% O_2 or less). Good regeneration occurs at temperatures of 1850°F or higher. Under these conditions, coal combustion efficiency of the present FBM/CBC overall system is less than 98%. Regeneration is aided by coal addition to the CBC. Sulfur balances are good when coal is fed to the CBC.

(6) At low bed sulfur levels, salt addition to the FBM acts primarily to increase combustion efficiency. When SO_2 capture is already 90% or better, the improvement in capture caused by salt addition is difficult to detect. Acting as a vapor-phase, homogeneous combustion catalyst, the amount of salt required is very small (less than 1% of coal weight). This enhancement is important, since higher carbon burnup in the FBM leads to a possible reduction in CBC cross-sectional area, and consequently lower CBC air rates, making higher SO_2 levels possible in the regeneration off-gas, at a given value of O_2.

Using the results of the tests in the FBM-CBC system, a regeneration section was designed and an FBM-CBC-REG three-vessel system built and tested to investigate the MK II concept. In the CBC, high O_2 yields high carbon burnup. In the REG (coal-fired), low O_2 yields high levels of bed regeneration and a low volumetric flow of high-SO_2 process gas (up to 10% SO_2). Various process arrangements are possible. FBM bed may be fed to the REG. REG bed may be flowed to the FBM or CBC or both. REG high-carbon fly ash may be fed to the CBC or discarded. Studies on the optimum arrangement were not performed.

1.4 Designs and Cost Estimates

This report contains a section on boiler design concepts and costs estimates prepared in 1971. Inflation since that time makes the absolute values presented here obslete However, cost studies done in 1973 show that the cost of a fluidized-bed boiler relative to the alternative furnace design is still favorable.

NOMENCLATURE

CBC	Experimental Carbon-Burnup Cell
$\bar{d}$	Weight mean particle diameter
FBC	Fluidized-Bed Column
K_{eq}	Equilibrium constant
ppm	Parts per million, volume basis
T	Bed temperature, °F
μm	Micrometers
1°	primary

2. CONCLUSIONS

Based on an analysis of the experimental work carried out under this program, the following conclusions were drawn:

a. A multicell fluidized-bed pilot scale boiler has been constructed which can operate at high air rates and achieve 99+% combustion efficiency. Because of the high air rate and combustion intensity, cost estimates for commercial boilers show that fluid-bed units will be less costly than conventional units.

b. The desired level of combustion efficiency may be achieved by recycling collected carbon-bearing fines from the primary combustor to a region of the fluidized-bed boiler called the Carbon Burnup Cell, in which the bed temperature is in the range 1950-2050°F and the O_2 is above 3%. The Carbon Burnup Cell requirements for bed depth, firing rate, and air rate have been determined. The tandem boiler-CBC operation was shown superior to fly ash reinjection to the primary cell, since carbon in fly ash is less reactive than coal.

c. Fuel costs are minimized; i.e., the system is optimized, when the boiler system is designed and operated so that approximately 90% of the fuel value, fed as coal to the fluidized-bed boiler, is consumed in the primary cells and 10% in the Carbon-Burnup Cell region.

d. This boiler has shown advantages in control of SO_2, NO_x, hydrocarbons, carbon monoxide and other pollutants.

e. Calcium sulfate formed in the low temperature (i.e., coal-burning) regions of the fluidized-bed boiler, either by use of particulate lime beds or by injection of fine limestone into inert ash beds, will not decompose in the high temperature Carbon-Burnup Cell if the residual oxygen level is maintained at above 3.5%, when no coal is added to the Carbon-Burnup Cell.

f. Decomposition of this calcium sulfate may be achieved by operation at low levels of residual oxygen, and/or by feeding coal to the CBC. However, the SO_2 concentration achievable with fine lime in a high velocity apparatus was relatively low, ~½%. SO_2 levels of 3 to 4% were achieved with the existing CBC using -8+20 mesh lime bed material but not with 3% O_2 in CBC flue gas. The existing prototype burnup cell is not optimized for simultaneous use as a regenerator, in the SO_2 Acceptor Process MK I configuration.

g. Injection of fine limestone does not appear to lend itself to a regenerative SO_2 control process for a second reason: no easy method of separating the regenerated fine lime from the burned-out coal ash is apparent; reinjection of coal ash to the boiler is undesirable.

h. Using lime beds in either the two-cell or three-cell systems, simultaneous achievement of 90% or better capture of SO_2 in the FBM and 98+% carbon burnup is relatively easy. Alternatively, 90% SO_2 capture and synthesis of a 3% SO_2 regenerator gas can simultaneously be achieved. But all three results: high capture, high burnup and high regenerator SO_2 concentration simultaneously required use of the three-cell system during this program. At the gas velocities in use, simultaneous achievement of all three was realized using a three-reactor system in which the high temperature functions (carbon burnup and SO_2 regeneration) occur in separate zones at different O_2 levels (the SO_2 Acceptor Process, MK II). A Carbon-Burnup Cell could be built using lower gas velocity and larger bed area in which the simultaneous 98% burnup-3% SO_2 synthesis conditions could be met. Alternatively, the Carbon-Burnup Cell could operate with a relatively deep bed and achieve the same results. It is believed that a two-cell system may be technically feasible, although experimental proof is required.

i. Using the three-cell SO_2 Acceptor Process, MK II regenerator flue gas containing up to 10% SO_2 by volume was achieved. Such a flue gas is believed suitable for economical sulfur recovery, H_2SO_4 synthesis or lime scrubbing treatment.

3. RECOMMENDATIONS

Near term optimization of the three-cell FBM/CBC/REG steam generating system is recommended. The regenerator would operate at low excess O_2 levels and return active bed material to the boiler. It is presently sized to yield a flue gas containing 5% or more SO_2 by volume*. It is not anticipated that the steady state SO_2 level in the regenerator flue gas will exceed 10%, although the upper limit value is not firmly known at this time. The FBM typically captures 90 to 95% of the sulfur released by coal combustion, using an active lime bed. The tandem FBM/CBC system will combust 98+% of the carbon fed in coal, using burnup techniques already established.

In order to realize the air pollution control potential of the multicell fluidized-bed boiler as rapidly as possible, the following further actions are recommended:

a. A set of cost and performance goals for fluidized-bed boilers should be established. This would include stringent air pollution control goals on SO_x, NO_x, hydrocarbons, CO, halogens, particulates, and plume opacity.

b. Pope, Evans and Robbins, together with one of the major boiler manufacturers and a public utility, should perform a detailed engineering design for a large coal-fired, multicell, fluidized-bed boiler which may meet these goals.

c. Based on the questions which arise about that particular design and related designs, an experimental program should be conducted to answer those questions.

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* Operation at over 10% SO_2 is not visualized.

d. If, based on the assessment of an actual design, EPA's goals will be met, a prototype boiler should be constructed. A unit capable of producing 10 to 40 MW(e) would be adequate to provide the needed operating experience and design data for much larger units.

e. Pending a decision to proceed with the rational plan outlined above, the following experimental work on the air pollution control aspects of the multicell, fluidized-bed boiler should be carried out:

(1) Studies are required to optimize the regeneration section of the multicell, fluidized-bed boiler.

(2) Tests of the SO_2 Acceptor Process should be conducted which will determine, for geographically matched limestones and coals, and various coal sizes, the required sorbent circulation and makeup rates.

(3) Experiments to indicate the directions in which reduced emissions of oxides of nitrogen may be achieved should be initiated.

(4) A definition of the particulate and plume opacity control requirements of a multicell, fluidized-bed boiler should be found in cooperation with a leading manufacturer of dust collection apparatus.

4. INTRODUCTION

4.1 Description of a Fluidized-Bed Boiler

4.1.1 The Need for a New Form of Combustion

Since 1962 PER has carried out design and experimental studies aimed at reducing the cost of utilizing coal as a boiler fuel, initially under the sponsorship of the Office of Coal Research, Department of the Interior, and later under EPA sponsorship as well.

These studies were concerned primarily with improvements in the design of plants and boilers for industrial steam generation; i.e., systems which would be large enough to supply power and/or process heat to a factory, but too large to be used by a laundry or apartment house, and too small to be used by a large electric utility.

Boilers in this size range were selected for development for a number of reasons which can be summarized by a single statement -- a novel boiler could be commercially successful in this size range with less development expense. Success at this level could then lead to scaling both downward and upward.

Conventional methods of firing coal, as a fixed-bed on a stoker grate, or as a suspension flame in a pulverized fuel burner, were not found to hold promise for major reductions in size and cost, regardless of development effort.

It was found that the high cost of a coal-fired industrial boiler compared to comparable oil or gas-fired boilers was primarily due to differences in furnace size. Oil and gas could be burned in a smaller furnace than could coal. For industrial-sized boilers, this difference in furnace size meant that oil and gas-fired boilers could be assembled in a factory and shipped to the user's site by rail, while coal-fired boilers of equivalent capacity could not. The need arose for a method of firing coal which would reduce the size of the furnace so that higher capacity coal-fired boilers could also be factory-assembled with consequent cost saving, compared to on-site construction.

An evaluation of alternatives led to the selection of the fluidized-bed boiler as the most promising method of achieving the goals of our sponsor. Later, the air pollution control potential interested EPA.

4.1.2 Defining a Fluidized-Bed Boiler

A fluidized-bed boiler is defined as a system which meets all of the following criteria:

a. The system's primary function is the generation of steam. Therefore, the materials of construction, the mode of operation, the arrangement, auxiliary power requirements, etc., are consistent with existing practices and economics in the conventional boiler field.

b. The fuel is added to and burned within a turbulent, aerated bed which has been termed a "fluidized bed."

c. A significant fraction of the heat released by the burning fuel is immediately extracted by heat transfer surface in contact with the turbulent bed.

A fluidized bed, in turn, is defined as a mass of particulate solids held in suspension by an upward current of fluid such that the bed has zero-angle of repose and exhibits certain other properties of a liquid. Among the liquidlike properties of a fluidized-bed which are important to the boiler designer, is that the bed becomes well-mixed, with sufficient agitation, and the bed material can be caused to flow about or out of the system without the aid of mechanical devices.

4.1.3 What a Fluidized-Bed Boiler is Not

A number of systems have been conceived in which partial or complete combustion is carried out within a fluidized bed, the most successful being the regeneration section of the fluid catalytic cracker developed for the petroleum industry. However, unless the system meets each of the criteria listed above, it would not be classified as a fluidized-bed boiler.

A fluidized-bed is not an exotic new system requiring the establishment of a new industry such as required for nuclear power development. The boilers will be built by the companies now making conventional boilers, in their existing shops. A few months of experience will provide a skilled boiler operator with a sufficient understanding of the new form of combustion to perform his job.

4.1.4 A Simplified Description of a Fluidized-Bed Boiler

A fluidized-bed boiler consists, in its simplest form, of an enclosure containing both boiler tubes and a bed of granular solids. The bottom of the enclosure is perforated, and air is forced into the enclosure to fluidize the solids and react with coal which will be added to the bed. Such a system is shown schematically in Figure 1. When the temperature of the bed is raised by an auxiliary means to above about 800^{o}F, bituminous coal added to the bed will ignite*. The temperature

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* The ignition temperature of anthracite was shown to be well above 800^{o}F.

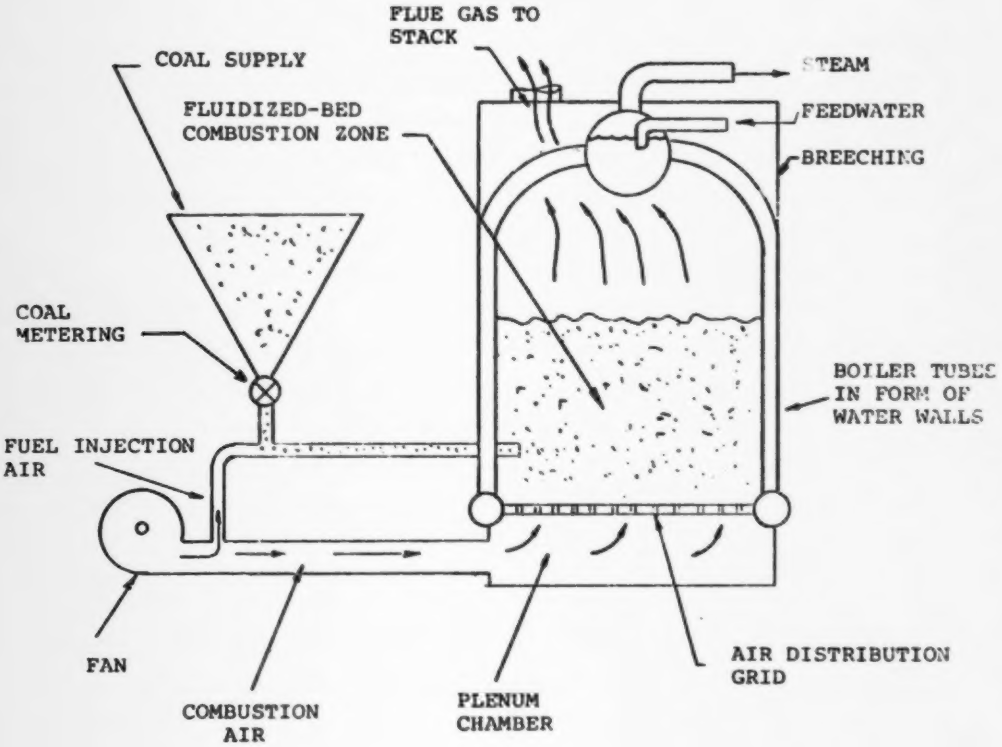
FLUE GAS TO STACK
COAL SUPPLY
STEAM
FLUIDIZED-BED COMBUSTION ZONE
FEEDWATER
BREECHING
COAL METERING
BOILER TUBES IN FORM OF WATER WALLS
FUEL INJECTION AIR
FAN
COMBUSTION AIR
PLENUM CHAMBER
AIR DISTRIBUTION GRID

FIGURE 1. SCHEMATIC OF FLUIDIZED-BED BOILER

will then rise until the system achieves a thermal equilibrium; i.e., the energy added to the bed by the burning fuel precisely equals the energy extracted by the boiler tubes touching and viewing the bed and the gases and dusts leaving the bed. At steady state, the fluidized bed consists almost entirely of inert particles with a small quantity of reacting coal.

This equilibrium temperature may be as low as 1200°F or as high as 2500°F, although a narrower range, 1450°F to 2050°F, is of practical interest. Below about 1450°F, it is difficult to completely consume carbon monoxide, while above 2050°F particles of certain fusible bed materials may couple and the bed collapse.

The depth of the fluidized bed may also have a very wide range -- from a few inches to many feet. Again, however, practical atmospheric systems must operate in the range of 12 to 48 inches. Below about 12 inches, the combustion efficiency degrades; above about 48 inches, the power required by the blowers becomes excessive*.

In specifying the operating temperature and bed depth, the energy release rate still remains unspecified. Rather than specify an energy release rate, however, it is more appropriate to specify an easily measurable equivalent, the mass flow rate of combustion air. The air rate may vary over the range of 100 lb/hr ft^2 of bed plan area to well over 1000 lb/hr ft^2 of bed plan area. Practical systems, operating at atmospheric pressure, will operate in the range of 500 to 900 lb/hr ft^2. The coal mass rate is then about 50-60 lb/hr ft^2.

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* Fluidized-bed boilers may be incorporated into combined cycles in which the hot flue gas at high pressure is passed through a turbine which is used to compress the combustion air as well as drive an alternator. In this system the economics are not so sensitive to bed depth. The pressure cycle leads to important size reduction possibilities. The reduced size of fluidized-bed boilers compared to pulverized fuel furnaces is an important advantage in pressure vessel design. Availability of turbines designed to run on fly ash remains limited, however.

From a specification of the air rate, the fuel rate, the bed depth, and bed temperature, it is possible to compute the apparent volumetric heat release rate, the superficial gas velocity and residence time and other parameters of interest. The air rate selected determines the size and density distribution of the particles which will make up the fluidized bed. Particles above a certain maximum size and density will sink to the air distributor. Particles below a certain minimum size and density will tend to be entrained in the gases leaving the bed and be carried out of the system*. The fluidized bed, in addition to being an averaging device, naturally selects those particles which it wishes to retain and rejects those whose properties fall outside the desired range. The particles which make up the bed may be any non-combustible, granular solid which is sufficiently tough** to retain its shape and size in a bed over an extended period. In some coals, the adventitious or "non-inherent" mineral matter leaves an ash which meets these criteria and the coal can be burned in a fluidized-bed composed of its own ash. For many coals, however, the ash is too fine to be retained in a fluidized bed unless it is deliberately sintered.

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* It appears characteristic of beds which are not monodisperse that fines added to the bed do not elutriate immediately but rather a pulse of fines is emitted with a decay curve corresponding to a zero order process.

** Methods of quantatively specifying "toughness" were beyond the program scope.

For these coals, a material must be added with the coal to make up for particles of starter bed which are lost to the system. This material, besides meeting the properties listed above, must also be inexpensive. In many areas, limestone may be the material of choice, because of its cost and relatively low density after calcination, which permits a relatively deep bed to be used with a moderate pressure drop through the bed. Limestone also possesses other properties of interest to the designer of a fluidized-bed boiler and these are discussed below.

4.2 Air Pollution Control Potential of Fluidized-Bed Boilers

4.2.1 What is Air Pollution?

The question is not facetious, for it has been seriously suggested that all the products of combustion of a fossil fuel constitute air pollution. Disasters have been predicted because carbon dioxide may increase to levels at which the earth's solar energy balance and global ecology are disturbed. However, the CO_2 absorption capacity of seawater and vegetation appear unlimited. While water-vapor release into the lower levels of the atmosphere has not been judged to be dangerous, the steam plume which is visible atop many stacks on cold, humid days arouses numerous calls to a community's air quality wardens.

The nitrogen which leaves the stack of a large power station at temperatures above ambient may affect the micrometeorology of the plant site. It is of value to sailplane enthusiasts. However, most responsible authorities would not classify carbon dioxide, water vapor, and warm nitrogen as pollutants. Classified (by law) as pollutants are products of incomplete combustion (carbon monoxide and a variety of hydrocarbons), gaseous oxides of sulfur and nitrogen, gaseous halides, and particulates. To these might also be added heavy metal vapors and natural radioactive isotopes in the particulate (e.g., K^{40}).

4.2.2 Pollution Control Potential of Fluidized-Bed Boilers

4.2.2.1 Discussion

4.2.2.1.1 General

Fluidized-bed combustion of coal is, in itself, not remedy of any of the pollutants. A fluidized-bed boiler is not *per se* a pollution control device. However, certain properties of fluidized-bed combustion and of a properly designed fluidized-bed boiler can be exploited to produce a steam supply which is "clean" at less cost than available alternatives.

By careful design, partial combustion products can be consumed within the fluidized bed and in the freeboard. Although no work appears to have been done on methods of reducing CO and C_xH_y (beyond decreasing the coal: air ratio), emissions on the order of 200 ppm or less for each can be anticipated. Additional research possibly a search for low cost combustion catalysts, may provide methods for a further reduction in emissions. Such an approach may be economical in a fluidized-bed boiler, but is less likely to be practical for conventional boilers.

4.2.2.1.2 Sulfur Oxides

By the use of an attrition-resistant limestone* as the bed material, it is possible to absorb virtually all the sulfur released by the coal. To achieve very low levels of emission, a very large excess of active lime is required in the bed and freeboard at all times*. Economics, in turn, require that the limestone be kept active by continuously stripping off much of the sulfur in a separate zone so as to regenerate the limestone. The off-gas of the regenerator can then be processed to recover sulfur or sulfur products or be scrubbed, using a portion of the lime produced by the boiler, or CBC fly ash which is naturally rich in lime. The scrubber effluent containing $CaSO_4$ must then be disposed of. Elemental sulfur is difficult to obtain by the use of alkaline earths as the principal sulfur acceptor. The use of alkali-based sorbents may prove more effective. The presence of less excess sorbent may be feasible and elemental suflur may be more readily obtained.

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* Some limestones have been found to possess inadequate attrition resistance. Note that hardness and attrition resistance are not synonymous.

The property of a fluidized-bed boiler which makes effective in situ sulfur control possible is the relatively low temperature (about 1500°F) of the medium in which the combustion occurs. The combustion bed may be kept cool by immersing heat exchange surfaces therein, or, alternatively, by circulating the sorbent bed through a separate heat exchanger. In addition, the refluxing

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* Pressurized boiler operation aids SO_2 capture but poses additional regeneration problems. For example, pressure favors formation of other calcium-sulfur compounds:

$$CaSO_4 + SO_3 \rightarrow CaS_2O_7$$

$$CaSO_4 + SO_2 \rightarrow CaS_2O_6$$

Thermochemical data are lacking for the pyrosulfate and dithionate type materials. Attempts to regenerate bed material may form gaseous sulfur oxides which then are resorbed by the reactions shown. Current difficulties in decomposing sulfated lime under pressure may be due to this mechanism.

of sorbent particles in the freeboard provides a degree of concurrency to an otherwise well-stirred system. Additional development to provide countercurrency in a multi-bed device (stacked beds) with bed regeneration may provide emission control of the order of 99+%.

4.2.2.1.3 Nitrogen Oxides

Nitrogen oxides are produced in any boiler through the fixation of atmospheric nitrogen and oxidation of nitrogenous compounds in the coal*. Because of the relatively low temperature of the combustion medium (∿1450-1600^{o}F), it had been anticipated from equilibrium data that nitrogen oxides from a fluidized-bed boiler would be very low when compared to conventional boilers. However, this has not been found to be the case. Measured values of NO_x from some benchscale fluidized-bed combustors have approached levels reported for the "worst conventional boilers"; measurements of NO in which tend to be on the low side due to the reactivity of NO in sample systems.

While little work has been done on methods of reducing NO emissions from a fluidized-bed boiler, the techniques suggested for conventional boilers provide a starting point. One such technique is the recirculation of flue gas. While this method may lead to unstable combustion in the conventional boiler, this problem will not occur in a properly designed fluidized-bed boiler. Experiments we ran (2) in which less than the normal supply of oxygen was sent through the fluidized bed, reduced nitrogen oxide emissions to below 200 ppm.

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* Argonne National Laboratories, in a well-designed experiment utilizing an "artificial air" where argon was substituted for nitrogen, demonstrated that nitrogen oxides can arise from the nitrogenous compounds in the coal.

Catalysts which promote the reduction of nitric oxide to nitrogen in the presence of reducing gases might also be applied when a fluidized-bed heat exchanger follows a fluidized-bed combustor. A search for catalysts which can be produced from the mineral matter in coal is a promising area for research. A goal of 50 ppm for NO_x would be set for this approach.

4.2.2.1.4 Other

While the mineral matter in coal may prove useful in controlling nitrogen oxides, it is usually considered a source of pollution; i.e., particulates. Here the fluidized-bed boiler possesses certain inherent advantages over most conventional boilers. Since the coal is crushed rather than finely ground, much of the adventitious mineral matter remains in the bed and is not entrained in the gas. Particles which are entrained may be removed by a relatively inefficient and inexpensive mechanical dust collector, compared to the requirements imposed by pulverized coal firing

Inherent ash, on the other hand, may be very fine and will be entrained by the gas regardless of coal particle size. Some of this will pass through a mechanical collector and must be removed by an electrostatic precipitator. This final collection may be difficult when there is little sulfur trioxide in the gas, due to lime bed operation, since the electrical resistivity of most of the ash matter (silica, alumina and ferric oxide) is high at conventional collection temperatures. However, in a fluidized-bed boiler in which a Carbon-Burnup Cell is applied, the carbon content of the fly ash from the primary combustor (>40%) should act as a natural conditioning agent. This would allow efficient dust removal at conventionally low temperatures from the gas leaving the 1° cells.

The conventional measure of particulate emissions (as well as gaseous emissions) has been weight of pollutant per unit energy input. Emissions of 0.02 grains/standard cubic foot, the most stringent standard yet proposed, are the equivalent of 0.04 lb/10^6 Btu of fuel input. This should be achievable from a fluidized-bed boiler equipped with a high efficiency electrostatic precipitator.

A second standard for particulates is concerned with the opacity of the plumes issuing from the power plant's stacks. Standards now proposed will require that the emission be "invisible". While the opacity of the plume is related to the weight of particulates, it is not congruent with weight. A relatively opaque plume may result, for example, when firing conditions result in incomplete combustion causing micron-sized carbon particles to form from cracked hydrocarbons. Gas turbines and many incinerators smoke through this mechanism. This will not occur in a properly designed fluidized-bed boiler. A second source of light-scattering particles is sulfur trioxide droplets formed as the dew point is reached in the gas stream. When the bed of the fluidized-bed boiler contains limestone, the quantity of sulfur trioxide has been shown to be vanishingly small.

A third source of small particles is the inorganic fumes and smokes which result when mineral matter is volatilized during combustion. Because of the low temperature of the combustion medium in a fluidized-bed (~1500°F) boiler, some of the mineral matter may not volatilize, and a fraction of that volatilized out of the coal matrix will condense on the bed material and fly ash. As much as 99% of certain volatile species could be retained in a fluidized-bed operating at about 800°C.

Chlorine, fluorine, arsenic and selenium are four minor or trace constituents in coal which may be released in flue gas. Other trace constituents include sodium, potassium, phosphorous, titanium and manganese. While these have not yet been considered for control from coal-fired power stations, it may be anticipated that, as other pollutants are brought under control, attention may be turned to these. A fluidized-bed boiler can be designed to control even these pollutants. Chlorine and fluorine might be controlled by adding some potassium carbonate to a fluidized-bed combustor. Arsenic and selenium will be quantitatively removed by lime when present in the combustor bed, along with part of the chlorine and fluorine.

Coal, like seawater, contains every element in the periodic table below Z=92, and every natural isotope of these elements*. Since, on a worldwide basis, about 3×10^9 tons of coal are burned each year, tonnage quantities of even the trace constituents enter the atmosphere from the stacks of coal-burning power plants. Fortunately, the majority of these elements add little to the natural background levels and are of no concern.

Some elements are present in sufficiently high concentration and are sufficiently toxic to be of concern. Among these would be barium, cadmium, lead, and mercury**. The natural radioactive elements--radium for example--might be added to this list. It would be anticipated that a large fraction of the lead and uranium would be tied up by lime. Cadmium and mercury may be sufficiently volatile to appear in the flue gas, uncombined, and add measurably to the background concentration downwind of the plant stack. For each element, with the exception of mercury, the enormous surface area and low temperatures of a fluidized-bed boiler or heat exchanger may reduce the quantity of these elements which enter the atmosphere in dangerous forms. A systematic search for methods of removing cadmium and mercury from flue gas is indicated. Possibly the fluidized-bed boiler may prove useful in applying the method economically. Even if the undesirable elements are fixed in the ash, they may reappear in the biosphere if the methods of ash disposal are inadequate. There is no evidence today that ash from fluidized-bed boilers operating with lime beds will be utilized in ways that fix the soluble constituents*. No investigations have been made of the potential uses to which the coarse stream (impure lime) or the fines stream (coal ash containing partially sulfated lime) from a fluidized-bed boiler could be put.

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* Seawater now contains some of the higher elements such as plutonium.

** Up to 25 ppb Hg in bituminous coal.

However, because these streams have not been vitrified by exposure to high temperature, there is some hope that the ash of a fluidized-bed boiler may be a useful raw material for processing into commercial materials in which the impurities become permanently fixed.

4.2.2.2 Conclusions

It is seen that a fluidized-bed boiler may be designed to reduce air pollution to low levels. The pollution control potential follows from certain inherent characteristics: the low temperature of the bed, the low temperature of the combustion, the high carbon content of the 1° cell fly ash, the stability of the combustion, and the general properties of a fluidized-bed. While relatively little work has been done on the control of pollutants other than sulfur oxides, the development of a system in which pollution control and economical steam generation are simultaneously optimized appears to be an attainable goal.

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* e.g., $CaSeO_4$ solubility 9 parts per 100.

4.3 Pope, Evans and Robbins' Prior Work

4.3.1 General

Pope, Evans and Robbins, under contract with the Office of Research and Monitoring of the Environmental Protection Agency (successor to the National Air Pollution Control Administration, Department of Health, Education, and Welfare), has characterized emissions from a fluidized-bed boiler developed for the Office of Coal Research, Department of the Interior. Three reports have been prepared describing this work (References 1, 2, and 3).

The pollution control aspects of a fluidized-bed boiler had been considered as early as 1965, when it was discovered that if coal distribution were uniform, smokeless combustion was achieved at attractively low excess air levels. Early in 1966, a test was conducted in which dolomite was mixed with the coal, and the sulfur oxide control potential of the process was demonstrated. A literature and patent search revealed that this approach to the control of sulfur oxides from a coal-fired boiler was novel, though a patent had been granted earlier on the use of limestone in a shale-burning process.

4.3.2 Once-Through Limestone Injection

Based on this early work, the studies conducted for EPA first investigated the use of relatively coarse limestone and dolomite.

It was determined that coarse stones would not retain more than about 30% of the sulfur released by the coal at reasonable Ca/S ratios. A "sulfate shell" theory was invoked to explain this.

A dolomite identified as 1337 was found to decrepitate rapidly and would not be retained in the bed. A limestone, 1359, was retained in the bed but could not be fully converted to $CaSO_4$ if sufficient stone were added to significantly reduce sulfur oxide emissions.

An unexpected result was that the SO_2 capture rate decreased with increasing bed temperature. Bench scale studies by others using simulated flue gas mixtures had indicated more rapid capture of SO_2 at 1800°F than at 1600°F, yet the SO_2 capture process, which was then assumed kinetically limited as well as irreversible, performed far better at 1600°F than at 1800°F. It was then understood that proper simulation required combustion-generation of SO_2 within the sorption bed.

In order to enhance the process rate by increased surface area, the stone was finely ground. Despite the bed's inability to retain indefinitely particles under about 30 U.S. Standard Mesh*, the efficiency of sulfur capture increased with decreasing particle size, implying among other things that the finest particles calcine faster.

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* As noted earlier, economically feasible fluidized-bed boilers operate at air rates between about 400 and 1200 lb/hr per ft^2 of bed surface which at 1500°F corresponds to superficial gas velocity of 5.6 to 16.7 feet per second.

It was determined that about an 80% reduction in SO_2 emissions could be achieved by injecting fine 1359 limestone at Ca/S = 2.5. Raw stone and hydrate both performed equally well, while stone precalcined by the supplier performed poorly. The behavior of the precalcined stone was considered anamalous, since it would have been expected to be better; perhaps it was dead-burned.

4.3.3 Regenerative Limestone Process

Toward the end of the pollution characterization effort (May 1969), it was discovered that beds composed almost entirely of coarse limestone could be made to release the accumlated sulfur that had been retained in a batch operation by increasing the coal-feed rate so as to increase the bed temperature and decrease the oxygen content of the flue gas. A regenerative cycle was devised and a patent application prepared for the "SO_2 Acceptor Process."** This discovery was felt to provide an explanation for the anomalous temperature behavior noted earlier. Sulfur retention in a fluidized-bed combustor was now seen as a reversible process sensitive to temperature and oxygen partial pressure.* The reversibility at the temperatures of interest exceeded thermodynamic predictions and may depend upon contamination of the CaO by coal ash ingredients. Alternatively, localized reducing conditions might be responsible.

Pope, Evans and Robbins also demonstrated in a coal-fired, fluidized lime bed combustor that the injection of small amounts of sodium chloride greatly increases the sulfur removal capacity of the lime. We also determined in bench scale experiments that salt dissolves $CaSO_4$ at the temperature of the fluidized-bed combustor. This activity enhancement effect was thought to be due to the removal of the "sulfate shell" from the lime particles, increasing the activity of calcium oxide. See Appendix E for further background information.

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* It was not determined whether the apparent effect of oxygen was direct, or whether the effect of oxygen on the reducing gases present in the bed governed the retention-release of sulfur.

** U.S. Patent 3,717,700, February 20, 1973

4.3.4 Pollutants Other Than SO_2

Pollutants other than sulfur dioxide were characterized--nitrogen oxides, hydrocarbons, and particulates.

Oxides of nitrogen were found to be dependent on the coal: air ratio; i.e., nitrogen oxides would increase as the amount of oxygen remaining in the flue gas increased*. It was not determined if this resulted because more nitrogen was fixed, or because less nitric oxide was decomposed as the reducing gases in the bed were consumed. However, experiments in which some air was diverted from the base of the bed to a port above the bed resulted in a marked reduction in NO. This might indicate that the reducing gases in the bed were decomposing the NO. These results might also have indicated that conditions near the air distributor were governing. Tests with a variety of distributor designs were not conducted. Nitrogen oxides were not materially and reproducibly affected by the injection of fine limestone, although workers at Argonne have shown a reduction in NO with limestone injection (Reference 4). An increase in NO with limestone injection has also been predicted on the basis of experiments at Esso Research and Engineering (Reference 5).

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* A dependence of NO output on coal particle size has also been observed. This dependence on coal size tends to suggest that at least some NO is due to thermal fixation of N_2 from air because the large coal particles are known to be hotter then the rest of the bed particles.

NO levels of less than 400 ppm were found in PER's boiler. 275 ppm was typical.

Hydrocarbons and, presumably, carbon monoxide* were found to be sensitive to the quality of the fuel distribution and to the coal:air ratio. With between 3% and 4% oxygen remaining in the flue gas, the hydrocarbons were reduced to below 100 ppm. Hydrocarbons were not affected by fine limestone injection.

The major fraction of the mineral matter in the coal appears as fly ash. The larger particles of adventitious matter remain in the bed. The bed material itself adds some particulate matter to the flue gas and when fine limestone is injected, essentially all of this appears in the fly ash stream.

At the high dust loadings, when fine limestone injection was used, an inexpensive, low pressure-drop, mechanical collector proved remarkably efficient (∿90% removal).

The particles not removed by the collector were all smaller than 20 μm. These particles would have to be removed by an electrostatic precipitator, bag collector, or wet scrubber. Because of the high carbon content (∿40%), the resistivity of the FBM fly ash may be sufficiently low to permit efficient electrostatic collection at low gas temperatures.

There was no reason to believe, on the basis of tests performed, that a fluidized-bed boiler would be unable to comply with the most stringent regulations governing particulate emissions at equal or lower cost than any other boiler.

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* HC was recorded continuously. CO was determined intermittently by Orsat and was not over 0.1%, the limit of detection; no continuous record available.

4.3.5 The Carbon-Burnup Cell

4.3.5.1 Tests in the Fluidized-Bed Column

Development of staged combustion was required to reduce ash carbon content to the 0-10% range. The major portion of the effort to produce the design correlations of fly ash combustion tests was carried out in the Fluidized-Bed Column (FBC) which, during these tests, had a plan area of 0.86 square feet.

Fly ash was burned in this device over a wide range of conditions, summarized as follows:

Bed temperature:	1750°F to 2140°F
Bed depth (static):	10" to 22"
Air rate:	385 to 1000 lb/hr ft^2
Fuel rate:	48 to 350 lb/hr ft^2
Carbon concentration of fuel:	28% to 65%
Heat removal rate:	15% to 40% of heat release

The key result of the test program was this performance model, which may be used to predict the combustion efficiency in a Carbon-Burnup Cell from the parameters which were found to control performance.

This model equation is as follows:

Combustion efficiency, % = (1)
-13.78
+0.05193 (bed temperature, ^oF)
+0.03973 (air rate, lb/hr ft^2)
+0.3831 (static bed depth, inches)
-0.7514 (carbon feed rate, lb of carbon/hr ft^2)
-0.1638 (inert feed rate, lb of inert/hr ft^2)
+0.0020 (carbon feed rate x inert feed rate, lb^2/hr^2ft^4)

Tests of this model, using data obtained in runs not used in the model derivation, indicated that the model was not limited to simply reproducing itself but was a useful prediction tool.

The residual oxygen content of the flue gas could also be predicted, although somewhat less accurately than the combustion efficiency. This model equation is as follows:

$$
\begin{aligned}
&\text{Residual oxygen, \% =} &&(2)\\
&22.91\\
&-0.007353\ (\text{bed temperature, }^{\circ}\text{F})\\
&+0.01118\ \ (\text{air rate, lb/hr ft}^{2})\\
&-0.1390\ \ \ (\text{static bed height, inches})\\
&-0.1521\ \ \ (\text{carbon feed rate, lb carbon/hr ft}^{2})\\
&-0.0151\ \ \ (\text{inert feed rate, lb of inert/hr ft}^{2})\\
&+0.0002653\ (\text{carbon feed rate x inert feed rate,}\\
&\qquad\qquad\text{lb}^{2}\text{/hr}^{2}\text{ft}^{4})
\end{aligned}
$$

A parametric study was performed to determine the optimum split in duty between the primary cells and the Carbon-Burnup Cell. The lowest fuel costs are realized if the Carbon-Burnup Cell burns about 10% of the input fuel. This is equivalent to saying that the primary cells should operate with about 3% residual oxygen in the flue gas. This coincides with a good trade-off between higher hydrocarbon at low O_2, and lower thermal efficiency at high excess air.

When finally divided (-325 mesh) No. 1359 limestone was injected into the FBM in this series of tests, less sulfur was removed than in the tests described in Reference 2 (~70% at Ca/S = 2.5 compared to ~80% at Ca/S = 2.5 measured previously). The reasons for the less favorable performance are unknown but may have been the result of a different injector design.

When partially sulfated fine limestone entered the Carbon-Burnup Cell along with the carbon-bearing fly ash, its sulfur content could be released as SO_2 if the CBC operated at a temperature of $\sim 2000^{\circ}F$ with residual oxygen $\leq$ 2.0%. The highest SO_2 values, $\sim$5000 ppm, were measured when the residual oxygen level was $\sim$0.2%. On the other hand, when the residual oxygen in the Carbon-Burnup Cell was in the range 3.5 to 6.0%, the fine sulfate did not decompose at CBC temperatures. In fact, in one test at 6% O_2, the partially sulfated limestone appeared to still be reactive at a bed temperature of $1980^{\circ}F$. An alternative explanation for this result appears to involve dynamic exchange of sulfur*. An important result was that the sulfur oxide emission from a Carbon-Burnup Cell may be as low as 350 ppm.

When no sorbent was present, the sulfur in the fly ash would burn with an efficiency equal to that of the carbon, or greater.

The data gathered on nitric oxide emissions from a Carbon-Burnup Cell were correlated less accurately than other parameters, though the highest emissions were detected at the highest bed temperatures. A mean value of 539 ppm was measured for all fly ash combustion tests in which the bed temperature was above $1900^{\circ}F$**. No efforts were made to reduce nitric oxide emissions and this remains one of the major areas requiring research. It was noted that the addition of coal to the fly ash feed would reduce the NO level, suggesting one line of potential research.

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* More about this in Section 6.2, page 6-20.

** At typical boiler operating conditions, about one-third the total NO emission is CBC; two-thirds FBM.

Particulate emissions for a Carbon-Burnup Cell were found to decrease with increasing bed temperature, possibly through agglomeration of ash matter to the sintered ash bed particles.

Considering that in some tests the fuel was $\sim$65% ash, particulate emissions were remarkably low. The pollution control potential of retaining a large fraction of the ash in the bed of a Carbon-Burnup Cell is an avenue of research which should be explored.

Hydrocarbons and carbon monoxide emissions were essentially nil when operating in the Carbon-Burnup Cell mode.

4.3.5.2 Tests in the Fluidized-Bed Module

Some tests were conducted in the Fluidized-Bed Module (FBM), an actual boiler with a grate area of $\sim$9 ft^2.

A simulated Carbon-Burnup Cell was appended to the FBM to determine the problem areas in operating a fluidized-bed boiler with two distinct temperature regions. This device was designated the CBC. The low freeboard CBC design and was necessitated by the FBM steam drum. It was found that for the small CBC ($\sim$1.1 ft^2) an opening in the barrier between the two regions of just 2 square inches permitted the desired temperature difference to be achieved. An interchange rate of 12,000 pounds of bed material per hour per square foot of opening was estimated from two transient heating tests.

A new coal feeder design also added at the same time performed very poorly until a number of minor alterations corrected the problems. Coal feeding to a fluidized-bed boiler is an area which requires further development effort.

Two fly ash feeders were tested in the CBC. One design, termed a mushroom feeder, performed well, giving a relatively even fuel distribution.

Two bed particle "knockouts" were also tested in the CBC. These both consisted of cylinders arranged in a triangular array. Due to the inadequate freeboard, without a baffle screen, large quantities of bed material had been expelled from the CBC; with either baffle screen, the quantity of bed material carried over was small. When the baffle screen was made up of water-cooled tubes, an initial heat transfer coefficient of ~35 Btu/ft^{2}hr$^\circ$F was measured. Here the temperature refers to the fluidized-bed temperature. This was approximately twice the value that would be predicted from simple radiation and convection, indicating that an active heat transfer region exists above the dense phase of a high velocity (hence turbulent) fluidized-bed.

When all operational problems had been overcome, the FBM/CBC achieved an overall combustion efficiency of approximately 99%.

4.3.6 Recommendations Based on Prior Work

The program was ended with recommendations for further study of the most promising results:

(1) A study of the regenerative mode of utilizing a limestone sorbent. If the sorbent were circulated rapidly enough, it appears possible to achieve low emission levels from the acceptor region of the boiler and relatively high concentrations from the regenerator region, suitable as feed to a sulfur recovery operation.

(2) A study of reducing NO_x emissions by disturbing the oxygen gradients within the fluidized bed. Recirculation of flue gas, burning natural gas and coal together, and studying the effect of air distributor design were recommended.

(3) Modification of limestone's sulfur capture capability by additive injection.

The first and third suggestions formed the basis of the experimental work conducted in the period November 1970 through July 1971, which is discussed in this report.

4.4 Specific Objectives of This Work

As noted in Section 1, SUMMARY, the two primary objectives of this work were to (a) investigate the pollution control potential of the concept of coal combustion in a fluidized bed of lime particles, with continuous regeneration of the sorbent by circulating the partially reacted lime between the primary combustion bed and a connecting high temperature zone in which carbon bearing fly ash from the primary bed is burned (i.e., the 2-cell system); and (b) investigate the potential of salt injection into a fluidized-bed coal combustor as a means of increasing the sulfur sorption capacity of limestone, as well as combustion efficiency.

The major operating, design, and economic factors of interest in the "SO_2 Acceptor Process" included: investigations of interaction of superficial velocity, temperature, fly-ash carbon burnup, and SO_2 production; verification of better than 90% sulfur capture using a better than 4% sulfur coal; demonstration of 98+% carbon burnup in the system; demonstration of high (3-4%) SO_2 concentration in the gases off the regeneration zone; consideration of rate of makeup limestone required; verification that discard of used bed material is unnecessary for SO_2 capture maintenance; and superheat production from the burnup cell flue gas cooler. The approach taken to accomplish these objectives was to first conduct batch nonregenerative limestone bed and salt tests in the FBC on a small scale; then to arrange components and demonstrate the regenerative process in long duration tests on a larger scale in the FBM/CBC. It was not found practical in the limited time available to study the effect of CBC bed levels different than the FBM bed level; nor was it practical to vary CBC grate area to collectively optimize gas velocity, carbon burnup and regenerated SO_2 concentration. Testing was also to be performed in a three-reactor FBM/CBC/REG configuration: the SO_2 Acceptor Process MK II.

The effects of the burnup cell and regeneration functions on the design and economics of the PER modular fluidized-bed combustor in the 30MW and 300MW sizes were to be identified.

5. APPARATUS

5.1 Pilot Scale Combustor, FBC

The small scale nonregenerating tests were conducted in a pilot scale combustor, designated the FBC for "Fluidized-Bed Column." The FBC consists of a rectangular combustion space, 12" x 16", having an air distributor as shown in Figures 2 and 3. In operation, air at ambient temperature, compressed by two blowers in series, enters a plenum below the air distributor, and passes up through a grid of buttons (bubble caps) and into the combustion chamber where it fluidizes the bed material and provides the oxygen for combustion. Fuel, coal and/or fly ash is pneumatically injected through a port at the base of the bed.

The air distributor contains a matrix of grid buttons mounted in a mild steel plate. The buttons are 303 stainless steel and designed to direct the air slightly downward toward the grid plate. This initial downward flow tends to eliminate stagnant areas. A cross section of a typical button is shown in Figure 4. Obviously, this is an anti-sifting design.

To reduce the heat loss to the waterwalls of the column, and thereby to allow the study of bed temperature effects independently of bed height, the unit was insulated internally as shown in the partial cross section of Figure 5. The water-cooled hood (used in CBC simulation) was insulated in a similar manner. By this means, high bed temperatures, in the 1800°F to 2100°F range, can be achieved with relatively deep beds (10"-22"). By use of water-cooling surface in the form of bayonets, also shown in Figure 5, the temperature of the bed can be adjusted by raising or lowering bayonets. It was found in coal combustion tests that considerable carbon fouling

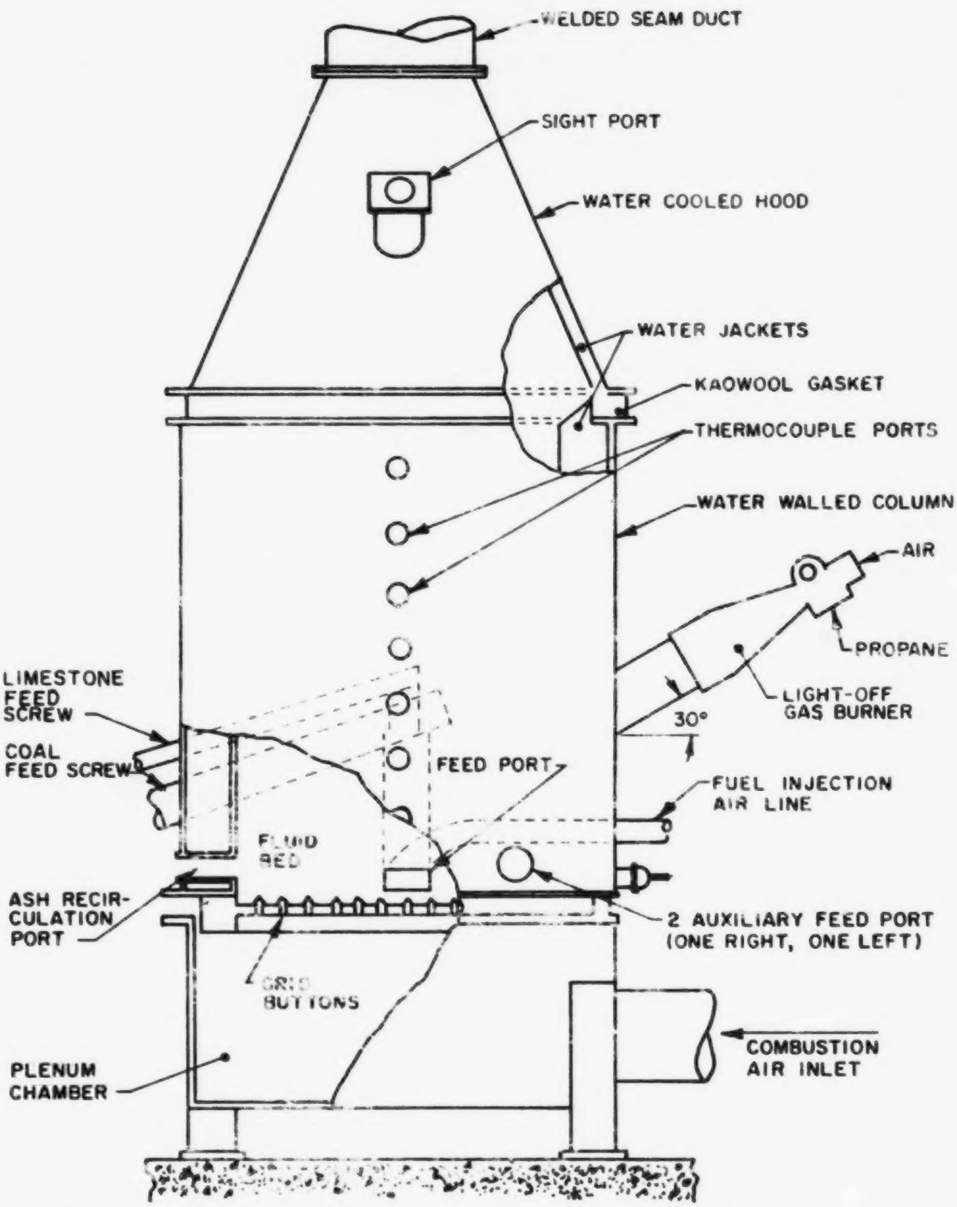
WELDED SEAM DUCT
SIGHT PORT
WATER COOLED HOOD
WATER JACKETS
KAOWOOL GASKET
THERMOCOUPLE PORTS
WATER WALLED COLUMN
AIR
PROPANE
LIGHT-OFF GAS BURNER
30°
LIMESTONE FEED SCREW
COAL FEED SCREW
FEED PORT
FUEL INJECTION AIR LINE
FLUID BED
ASH RECIRCULATION PORT
2 AUXILIARY FEED PORT (ONE RIGHT, ONE LEFT)
GRID BUTTONS
COMBUSTION AIR INLET
PLENUM CHAMBER

FIGURE 2. FLUIDIZED-BED COLUMN (FBC) CONSTRUCTION DETAIL FRONT VIEW

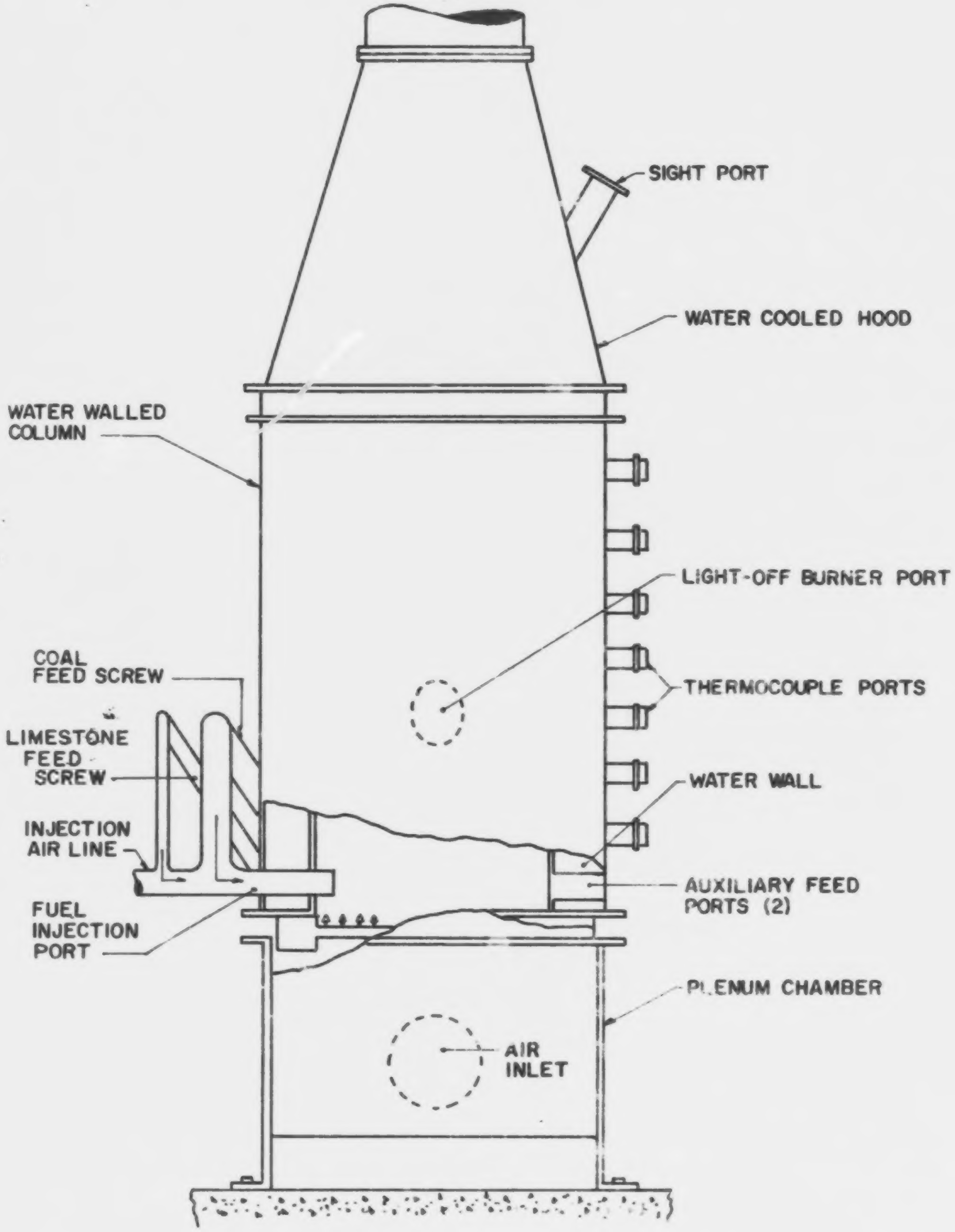
SIGHT PORT
WATER COOLED HOOD
WATER WALLED COLUMN
LIGHT-OFF BURNER PORT
COAL FEED SCREW
THERMOCOUPLE PORTS
LIMESTONE FEED SCREW
WATER WALL
INJECTION AIR LINE
AUXILIARY FEED PORTS (2)
FUEL INJECTION PORT
PLENUM CHAMBER
AIR INLET

FIGURE 3. FLUIDIZED-BED COLUMN (FBC) CONSTRUCTION DETAIL-SIDE VIEW

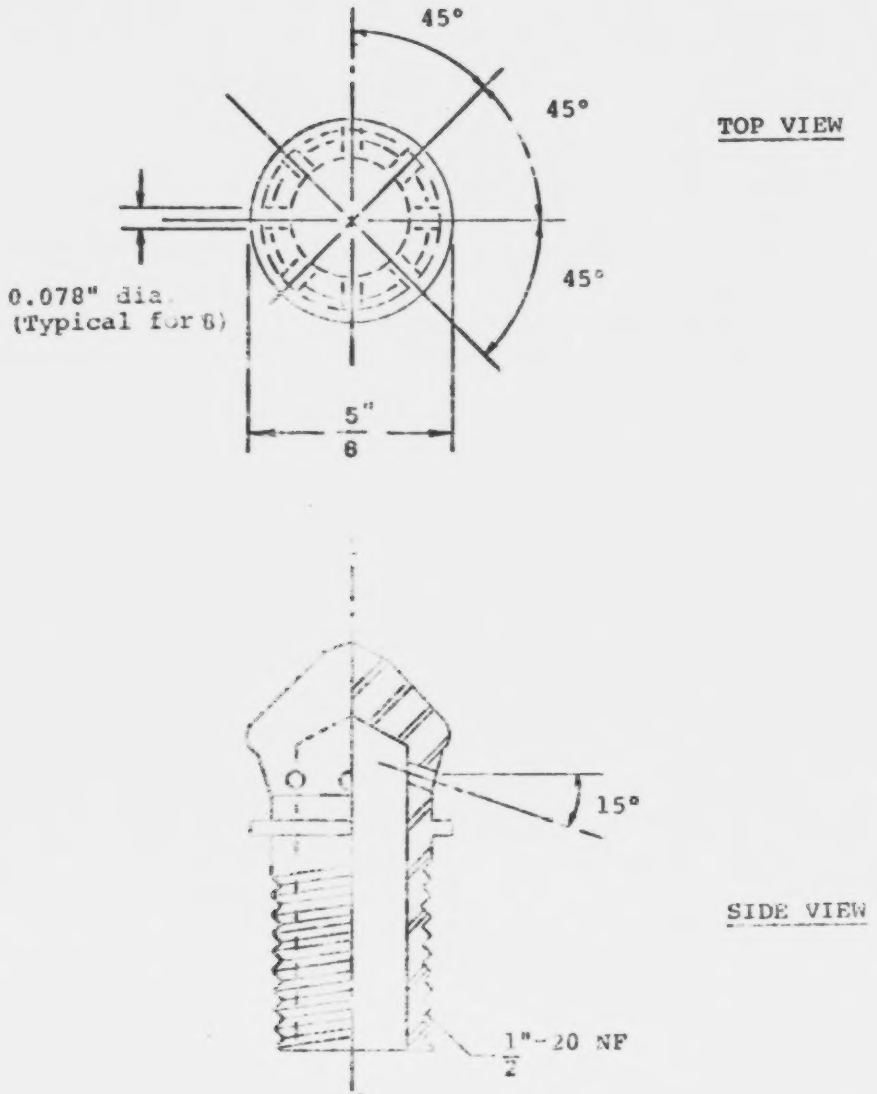
45°
45°
TOP VIEW
45°
0.078" dia.
(Typical for 8)
5"/8
15°
SIDE VIEW
1/2"-20 NF

FIGURE 4. AIR DISTRIBUTION GRID BUTTON

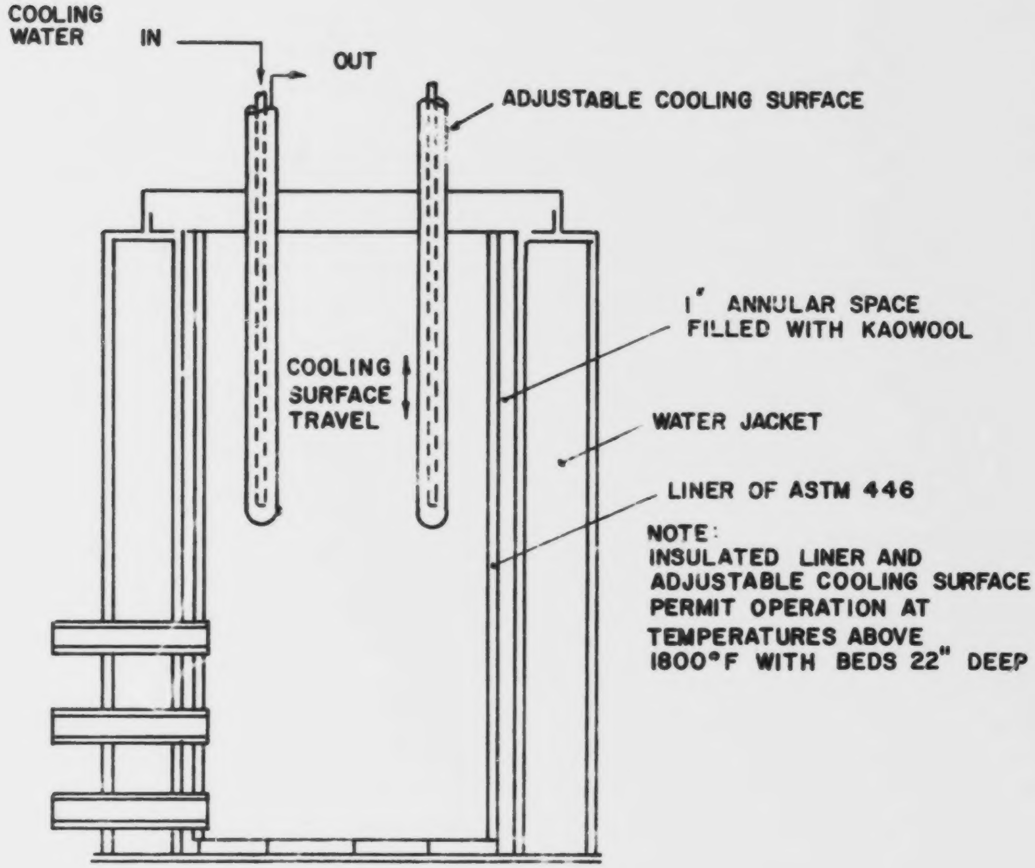
COOLING WATER
IN
OUT
ADJUSTABLE COOLING SURFACE
1" ANNULAR SPACE FILLED WITH KAOWOOL
COOLING SURFACE TRAVEL
WATER JACKET
LINER OF ASTM 446
NOTE:
INSULATED LINER AND ADJUSTABLE COOLING SURFACE PERMIT OPERATION AT TEMPERATURES ABOVE 1800°F WITH BEDS 22" DEEP

FIGURE 5. SECTION THROUGH FLUIDIZED-BED COLUMN SHOWING INSULATED STEEL LINER AND ADJUSTABLE COOLING SURFACE

of the water-cooled bayonets occurred, due to condensation and pyrolysis of coal volatile constituents. This did not occur when the fuel was carbon-bearing fly ash.

The insulation consists of a sleeve or liner of ASTM 446, one of the most refractory steels, backed with 1" of Kaowool, a refractory insulation. With the insulated sleeve, the FBC has an internal cross section of 0.86 ft^2.

The bed material consisted of 1359 limestone crushed and screened to -8+20 U.S. Standard sieve size.

The bed is heated to coal ignition temperatures with a premix gas burner flame directed downward into the bed, as shown in Figure 2. The ignition procedure involves fluidizing the bed material with minimum air flow, raising the bed temperature to 800°F, and then injecting coal until the combustion is self-sustaining. About ten minutes are required for the ignition procedure.

The bed temperature is monitored with several thermocouples spaced vertically in the combustor. Kaowool seals were provided to prevent flue gas leakage out of the system. Specifications for the FBC are presented in Appendix A.

The fuel feed system is capable of delivering $\sim 3.5 \times 10^6$ Btu/hr. The air feed system is capable of delivering oxygen sufficient for a heat release of $\sim 2 \times 10^6$ Btu/hr. These are both in excess of the actual operational values.

The FBC test system is shown schematically in Figure 6. Combustion products from the FBC pass through a heavy gauge welded seam duct, through an optional induced draft fan*, through a dust collector and on to

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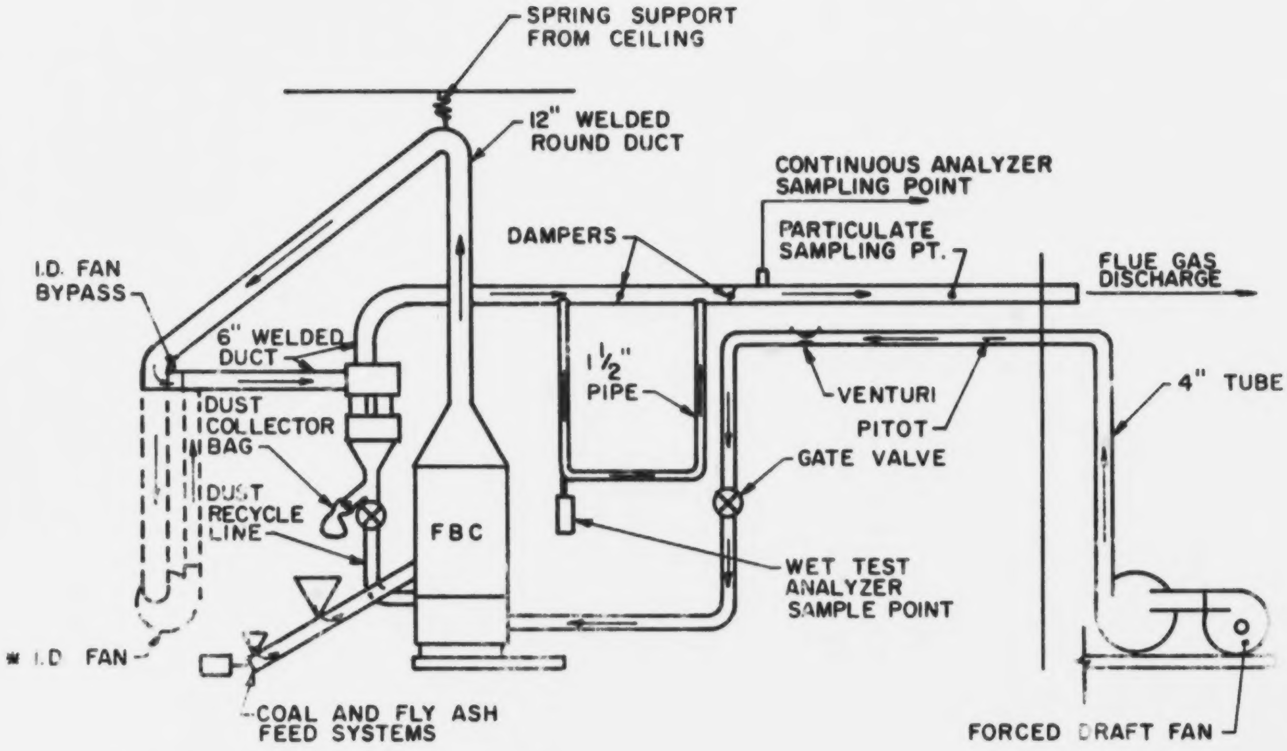
* The induced draft fan was not used in this test series.


SPRING SUPPORT FROM CEILING
12" WELDED ROUND DUCT
CONTINUOUS ANALYZER SAMPLING POINT
DAMPERS
PARTICULATE SAMPLING PT.
I.D. FAN BYPASS
FLUE GAS DISCHARGE
6" WELDED DUCT
1 1/2" PIPE
VENTURI
4" TUBE
DUST COLLECTOR BAG
PITOT
GATE VALVE
DUST RECYCLE LINE
FBC
WET TEST ANALYZER SAMPLE POINT
* I.D FAN
COAL AND FLY ASH FEED SYSTEMS
FORCED DRAFT FAN

* I.D. FAN NOT USED IN THIS TEST SERIES

FIGURE 6. SCHEMATIC OF FBC AIR AND EXHAUST GAS DUCTING SHOWING SAMPLING POINTS

the analytical system and stack. The slanted configuration of the duct between the FBC and dust collector provides gas cooling without causing wall surface temperatures to fall below the dew point of sulfur trioxide (~360°F). A control damper may be used to adjust back pressure on the system.

Fluidizing combustion air, provided by two series blowers located external to the test area, was monitored both by a pitot tube and a venturi meter located in the long entrance duct. A gate valve in the line was used to control air flow to the unit.

The fuel and additive feed rates were controlled by variable speed drives on the feed screws. The pneumatic fuel feed system was capable of feeding 250 lb/hr. Collected dust from the overhead cyclone was discharged into bags and weighed. A dust recirculation system, indicated in Figure 6, was available but was not used in this test series. Locations of thermocouples are described in Section 5.5 - Instrumentation. Salt feed into the pneumatic coal feed stream was by screw feeder or in some tests by metered aqueous solution injection. An isokinetic sampler is provided for studying stack particulate emissions.

5.2 Full-Scale Boiler Module, FBM

The full-scale boiler module, designated the FBM, is a boiler unit capable of generating steam under pressure up to 300 psig. In this unit, the fluidized bed is contained in a rectangular enclosure in which each wall is a row of vertical boiler tubes seal-welded so as to form a gas-tight enclosure. The FBM represented one half cell of the first multicell, fluidized-bed boiler concept developed under contract with the Office of Coal Research. Two modules placed back to back would comprise one cell. A number of cells placed side by side without intervening insulation would have made up a full-scale boiler.

A cut-away sketch of the FBM as it was at the beginning of the program is provided in Figure 7. A simulated Carbon-Burnup Cell (CBC), added at the rear of the FBM, is also shown in Figure 7. The CBC is discussed later. The FBM cross section is ∿18 x 72 inches, 9 square feet, roughly seven times the unsleeved FBC cross section. The bed is surrounded by vertical boiler tubes which extend from two cross headers below the grid plate to the steam drum. No other tubes are placed in the bed. The boiler tubes are joined together by welded fins and are backed by insulation. The fins do not extend the full height so that flue gas passes between the tubes at the top of the unit and around the steam drum.

The combustion space is accessible through a water-cooled panel at the front of the unit. The panel contains a view port and a premix gas burner used to fire the bed. The burner directs a flame downward onto the front of the bed. Two pneumatic feed ports are provided below the access panel, one for an optional coal feed or flyash reinjection tube, and the other for the makeup bed material feed tube. Four optional pneumatic feed ports are provided at the bottom of one side of the FBM, shown in Figure 9. For this test series, a vertical, split coal feeder, shown on Figures 7 and 9, was utilized. Other feeder designs have been used in other programs.

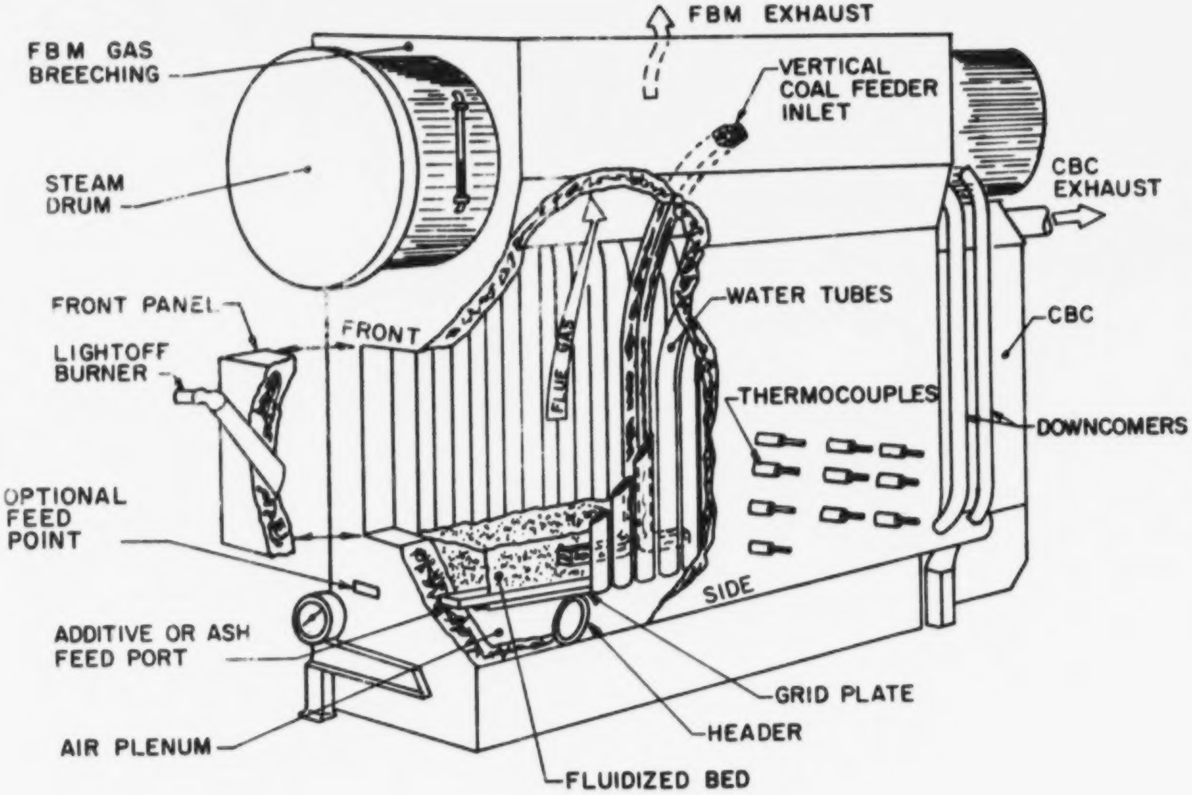

FBM EXHAUST
FBM GAS BREECHING
VERTICAL COAL FEEDER INLET
STEAM DRUM
CBC EXHAUST
FRONT PANEL
FRONT
FLUE GAS
WATER TUBES
CBC
LIGHTOFF BURNER
THERMOCOUPLES
DOWNCOMERS
OPTIONAL FEED POINT
SIDE
ADDITIVE OR ASH FEED PORT
GRID PLATE
HEADER
AIR PLENUM
FLUIDIZED BED

FIGURE 7. FLUIDIZED-BED MODULE (FBM) INTERNAL CONSTRUCTION
(INITIAL CONFIGURATION)

From the plenum at the base of the unit, air is directed upward through the grid and then through the bed. The grid consists of a mild steel plate containing buttons of the same spacing and design used in the FBC operation. The bed material used in the FBM tests was either 1359 limestone or sintered ash, double screened, -8+20 U.S. Standard mesh. The static bed depth may be varied from 6" to over 30", although the useful range is narrower. A bed sampling pipe and valve are provided. Thermocouples were mounted throughout the bed, as shown in Figure 8. Detailed specifications of the FBM are presented in Appendix B. Figures 9 and 10 show section views through the FBM and the two CBC's.

In operation, the bed is raised to the ignition point of coal by use of the gas burner. Combustion of the coal begins in the vicinity of the light-off burner flame and propagates rapidly throughout the bed. Firing with a coal input of 800 lb/hr, the FBM produces 200 psig steam at the rate of 5,000 lb/hr. The energy not absorbed by the waterwalls and steam drum leaves the FBM as hot products of combustion. Two water-cooled tube arrays to simulate the convection bank and economizer of a conventional boiler system were installed in the ducts beyond the FBM to absorb some of this energy.

A schematic drawing of the FBM test system is shown in Figure 11. Air from an external forced-draft fan passes through the air preheater and into the FBM plenum. Coal feed is controlled by the speed of a screw feeder which drops the coal into a pneumatic feed tube at the injection port*. Sorbent materials were screw fed to a pneumatic injection line at a rate controlled by a variable speed screw drive. Ash recirculation

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* When salt is fed to the FBM, it is screw fed from a weighing hopper and mixed with the coal feed.

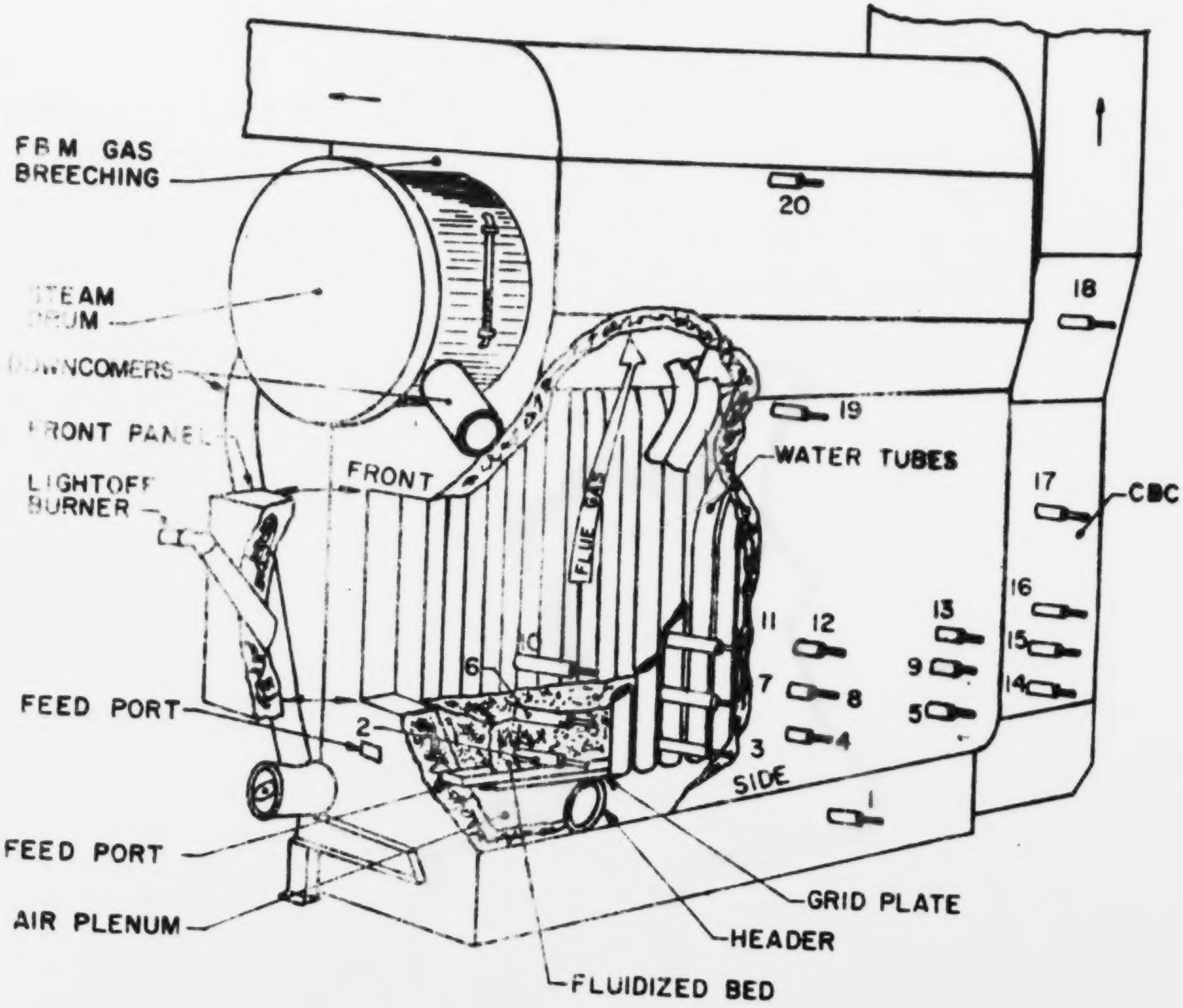
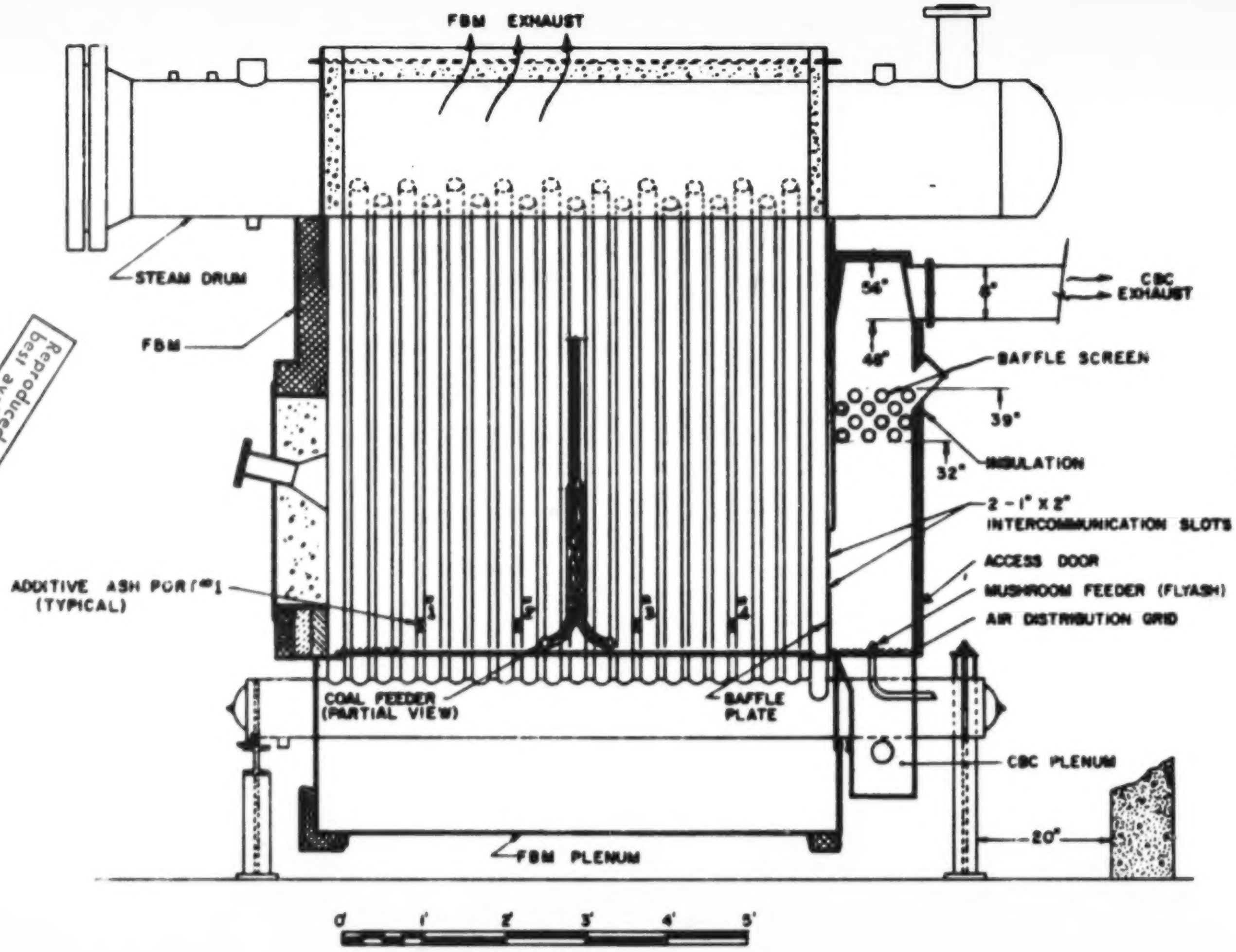

FBM GAS
BREECHING
STEAM
DRUM
DOWNCOMERS
FRONT PANEL
LIGHTOFF
BURNER
FEED PORT
FEED PORT
AIR PLENUM
FRONT
FLUE GAS
WATER TUBES
CBC
SIDE
GRID PLATE
HEADER
FLUIDIZED BED
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
17
18
19
20

FIGURE 8. FLUIDIZED-BED MODULE (FBM) INTERNAL CONSTRUCTION WITH TALL CBC SHOWING NUMBERED THERMOCOUPLE LOCATIONS


FBM EXHAUST
STEAM DRUM
FBM
ADDITIVE ASH PORT #1
(TYPICAL)
#1
#2
#3
#4
COAL FEEDER
(PARTIAL VIEW)
BAFFLE
PLATE
FBM PLENUM
CBC PLENUM
20"
54"
6"
48"
CBC
EXHAUST
BAFFLE SCREEN
39"
32"
INSULATION
2 - 1" X 2"
INTERCOMMUNICATION SLOTS
ACCESS DOOR
MUSHROOM FEEDER (FLYASH)
AIR DISTRIBUTION GRID
0' 1' 2' 3' 4' 5'
GRAPHIC SCALE

Reproduced from best available copy.

FIGURE 9. SECTION THROUGH FLUIDIZED-BED MODULE (FBM) AND CARBON-BURNUP CELL (CBC) (INITIAL CONFIGURATION)

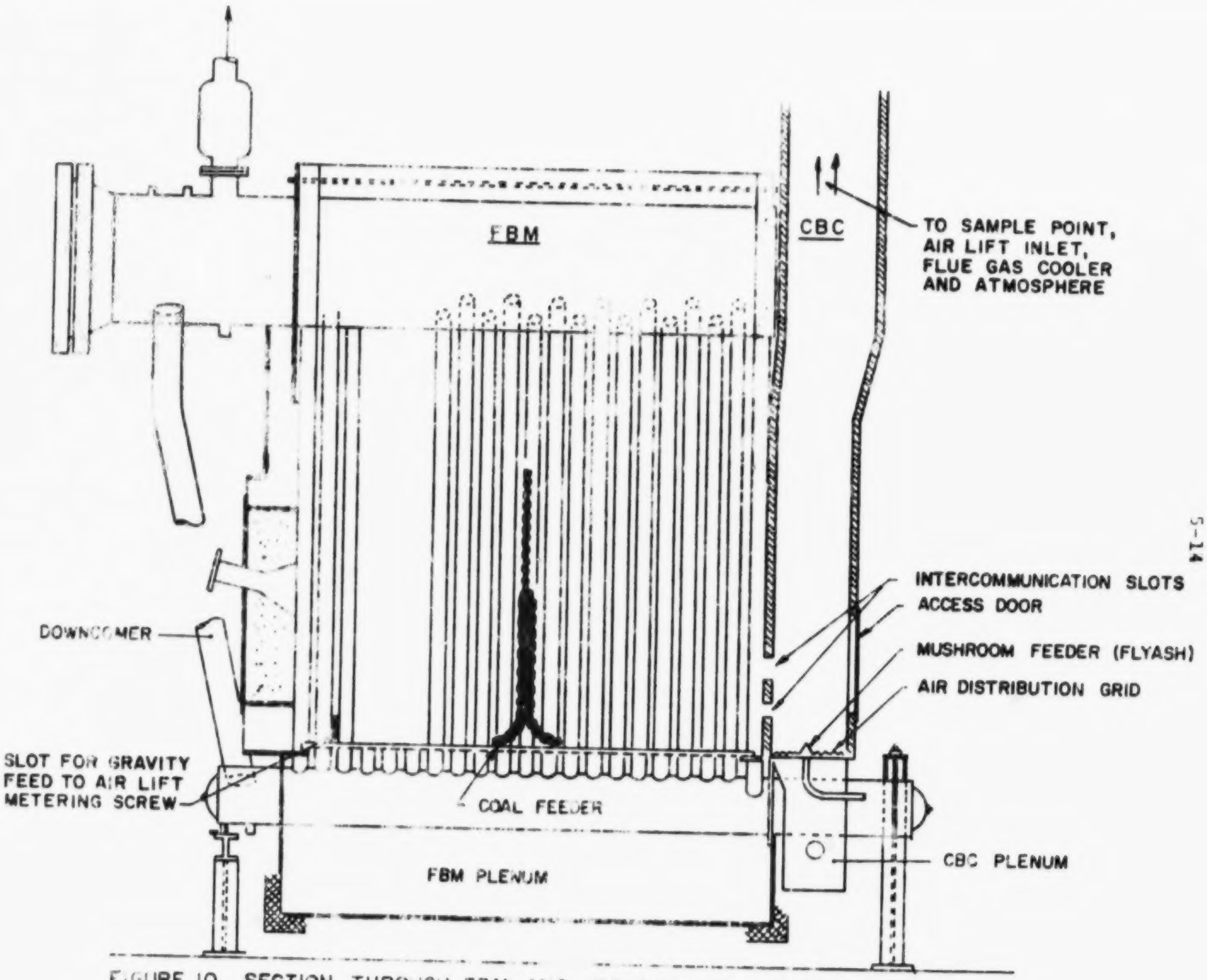

FBM
CBC
TO SAMPLE POINT, AIR LIFT INLET, FLUE GAS COOLER AND ATMOSPHERE
INTERCOMMUNICATION SLOTS
ACCESS DOOR
MUSHROOM FEEDER (FLYASH)
AIR DISTRIBUTION GRID
DOWNCOMER
SLOT FOR GRAVITY FEED TO AIR LIFT METERING SCREW
COAL FEEDER
FBM PLENUM
CBC PLENUM

FIGURE 10. SECTION THROUGH FBM AND CBC (REVISED CONFIGURATION)

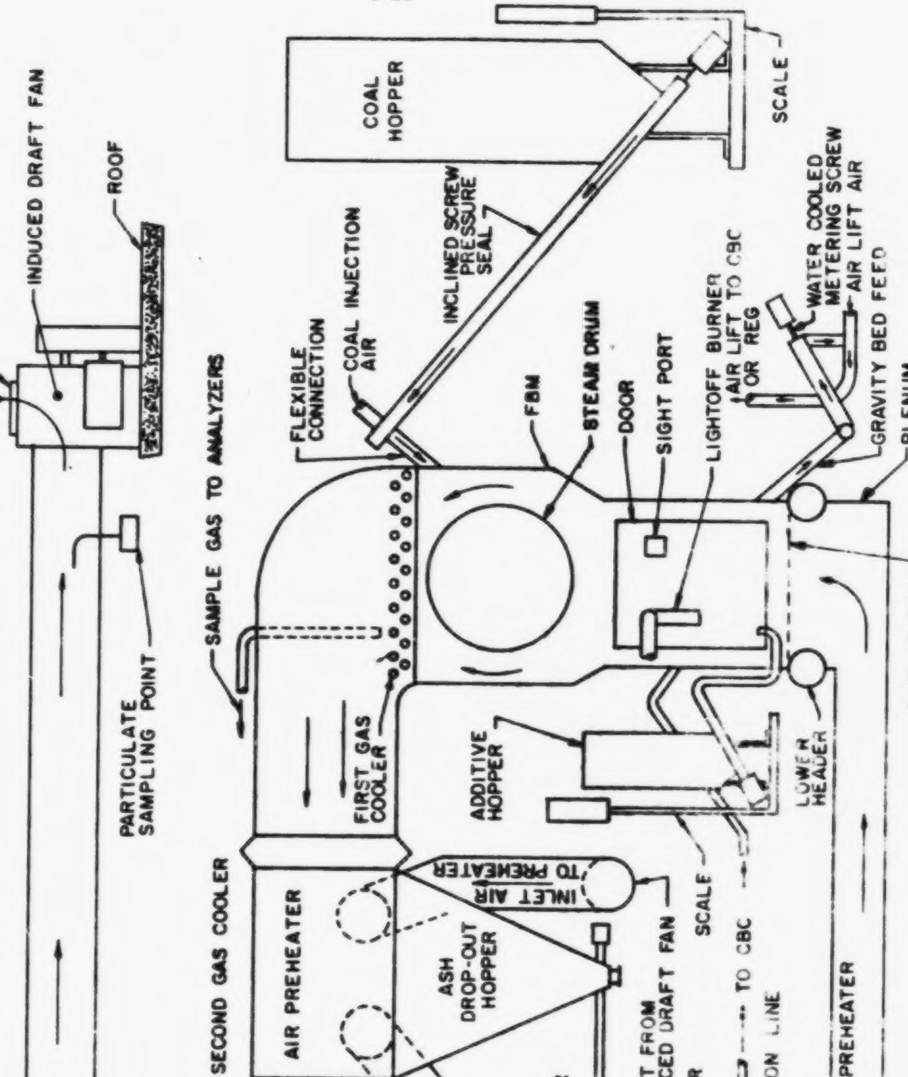

STUB STACK
INDUCED DRAFT FAN
ROOF
SAMPLE GAS TO ANALYZERS
PARTICULATE SAMPLING POINT
FLEXIBLE CONNECTION
COAL INJECTION AIR
INCLINED SCREW PRESSURE SEAL
COAL HOPPER
SCALE
FBM
STEAM DRUM
DOOR
SIGHT PORT
LIGHTOFF BURNER
AIR LIFT TO CBC OR REG
WATER COOLED METERING SCREW
AIR LIFT AIR
GRAVITY BED FEED
PLENUM
GRID PLATE
LOWER HEADER
ADDITIVE HOPPER
SCALE
FIRST GAS COOLER
INLET AIR TO PREHEATER
SECOND GAS COOLER
AIR PREHEATER
ASH DROP-OUT HOPPER
DUCT FROM FORCED DRAFT FAN
TO CBC
ASH REINJECTION LINE
SCREW FEEDER
STAR FEEDER
HOT AIR LINE
INLET AIR FROM PREHEATER
DRAFT BALANCE DAMPER
DUST COLLECTOR
DUST COLLECTOR HOPPER

FIGURE 11. SCHEMATIC OF FBM TEST SYSTEM SHOWING VARIOUS SUBSYSTEMS.

FIGURE 11. SCHEMATIC OF FBM TEST SYSTEM SHOWING VARIOUS SUBSYSTEMS.

is accomplished by pneumatic transport of fly ash from the dust collector through a star feeder. Bed material to be airlifted for regeneration is withdrawn from the FBM and passes through a variable speed,water-cooled screw feeder.

Flue gas from the FBM passes across the first gas cooler (convection bank) above the steam drum to reduce temperature before the gas enters the air preheater. As the flue gas passes through the air preheater, a portion (the coarse fraction) of the fly ash drops out and is collected in the hopper (see Figure 11). The bulk of the fly ash is removed by a multi-cone collector downstream of the air heater. During recirculation, the coarse fraction (ash knocked down by the preheater) is screw fed into the dust collector hopper. This unfortunately tends to cause inhomogeneous feed composition to the CBC. From the collector, the gas flows through a long duct to an induced draft fan and then to atmosphere. A damper is provided in the ducting to control pressure in the combustion chamber. The system is operable without the induced draft fan, but is not usually run pressurized. An isokinetic sampler is provided for the study of stack particulate emissions.

5.3 The Pilot Scale Carbon-Burnup Cell, CBC

The pilot-scale burnup cell, designated the CBC, was an appendage to the FBM It was to operate parallel with the FBM, at a higher temperature but with a common bed. The design of this system was based on the results of the previous phase of this program (Reference 1). The initial CBC consisted of an insulated rectangular box fabricated of ASTM 446, a steel with a relatively low coefficient of expansion capable of resisting oxidation at 2000°F, but of low ductility.

Although the internal dimensions of the CBC reactor zone were constant throughout the test program, several detail modifications were required before the CBC performed its functions of burning fly ash efficiently and with low bed loss; the system was developmental.

5.3.1 The Initial CBC

The initial CBC, shown in Figures 7 and 9, was that used in the previous test program (Reference 1); the CBC, after modification, is shown in Figures 8 and 10. As shown in Figures 7 and 9, the CBC was added to the rear* of the FBM. Its air distributor, 10-5/8" x 15-5/8", or 1.13 square feet, was identical in design and at the same elevation as that of the FBM. In the initial CBC configuration, the distance from the top of the air distributor to the roof of the CBC was 56"**. Starting 48" above the air distributor, the exhaust was installed as an 8" I.D. horizontal duct. The duct extended horizontally under the FBM steam drum, through the building wall and then ran vertically to a dust collector installed on the laboratory roof. A gas sample outlet was provided in the duct. Dust collected here could be wasted, routed to the FBM via the four-port injector described earlier, or back into the CBC. For Run No. 171H, a weighing ash silo was provided. After a few hours of operation, this device became inoperative.

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* The front of a unit is designated as the side in which the access opening is located. The front of the FBM is at the left of Figure 9. The front of the CBC is at the right in Figure 7.

** The dimensions of the unit were set by the design features of the FBM and its location in the laboratory. The height was set by the location of the FBM steam drum. The width by the location of the downcomers (see Figure 7) and the depth by a structural wall 20" beyond the boiler supports as shown in Figure 9. Except for the height, the dimensions were appropriate for the expected firing rate, 500,000-750,000 Btu/hr.

Fuel for the CBC was fly ash generated by the FBM, plus its own fly ash and coal, if required, to maintain the temperature at the $1900^{\circ}F$-$2100^{\circ}F$ level. Although the first trial injector was similar to that of the FBC (a square, horizontal tube jutting into the unit), improved operation was obtained with a "mushroom" feeder as shown in Figure 9. The mushroom feeder, as the name implies, is a solid cone. Its underside might be described as a hemi-toroid but the underside of a mushroom with short stem is a suitable description. It is fixed by four solid connections to the air distributor. An open vertical pipe extends up through the plenum and air distributor and ends below the stem of the mushroom. A fuel/air suspension will leave the pipe and be deflected by the underside of the mushroom into an isotropic stream with horizontal inertia. While the mushroom had been developed in 1967 for coal feeding, it is most suitable for dry materials such as fly ash.

Fluidizing-combustion air is withdrawn after the air heater from the main FBM air supply*. The design of the mushroom feeder makes it capable of supplying a significant fraction of the fluidizing air although the plenum and air distribution grid were always used. The FBM and CBC interface at the rear of each unit. To provide for a common fluidized-bed, the fins between boiler tubes making up the FBM's rear wall were removed. Looking toward the rear of the FBM, the initial configuration had five slots, 18" high and 1" wide, a 2" O.D. boiler tube between each slot. The back of these tubes (in the CBC) was insulated by a semicircle of insulation held in place by a thin metal shield. For bed intercommunication, holes cut in the baffle provided the desired opening, which was 2 square inches in this program.

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* Provisions also exist for feeding unheated CBC air in the unusual event a higher than desired CBC temperature should develop.

Two designs for flue gas "particulate knockouts" were tested, based on information obtained in April 1970 from BCURA Industrial Laboratories under an agreement between the EPA and England's National Coal Board. A knockout was required because the limited height of the initial CBC (48" to horizontal exhaust) resulted in excessive carryover of bed particles. The knockout is shown in Figure 9, labeled "baffle screen".

The first design consisted of an irregular triangular array (4 rows) of 1" tubes connected by U-Bends. Water flowed into the tube at the bottom and front of the unit and out the tube at the top and front. Thermometers installed in the inlet and outlet permitted the temperature rise to be determined. A total of 5.2 square feet of heat exchange surface was provided by the screen. The effect of this surface was described in the previous report in this series (Reference 1).

The second screen design consisted of a similar array of uncooled rods fabricated of ASTM 446. This screen, as well as the first, was installed within the CBC extending from the 32" level to the 39" level.

The test procedures for both the FBC and FBM/CBC operations involved igniting the bed and stabilizing the combustion at the desired bed temperature until steady-state conditions prevailed. Steady state was assumed when the bed temperature was constant and the gas composition detectors indicated constant values. A bed sampling pipe and valve are provided.

5.3.2 The Modified CBC

As development proceeded, it became obvious that the low CBC freeboard was placing intolerable constraints on system operation; especially carbon burnup efficiency and bed carryover. Therefore, a revised CBC* was built and the FBM steam drum and downcomers modified to allow a vertical CBC flue configuration (Figures 8 and 10). The modified CBC cross-section is a refractory cement box within a 446 steel shell. A flue gas cooler is incorporated in the top of the CBC. Sample gas is withdrawn below the cooler. FBM bed material may be airlifted into the CBC at a point above the gas sampler. The water-cooled tubes then act as a knockout for air lifted bed material entrained in the CBC flue gas rather than CBC entrained bed particles. When FBM bed material was airlifted, it was withdrawn at a point remote from the CBC and its rate was metered by a variable speed, screw feeder which was water-cooled**. By use of refractory insulation, improved heat economy is achieved and CBC coal feed should not be necessary to maintain the desired temperatures.

5.4 The Pilot Scale Bed Regenerator, REG

Following testing of the two-cell SO_2 Acceptor Process (in which the CBC is employed as the sorbent regenerator), it appeared desirable to physically separate the high temperature functions, Carbon-Burnup Cell (high O_2 with fly ash feed) and sorbent regeneration (low O_2 with coal feed). The ORM Project Monitor accordingly modified the contract work statement.

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* Also of 1.1 ft^2 grid area.

** It is important to note that a fluidized-bed boiler system using airlift bed transfer contains the basis for an oversize rejection classifier, permitting large lump coal utilization.

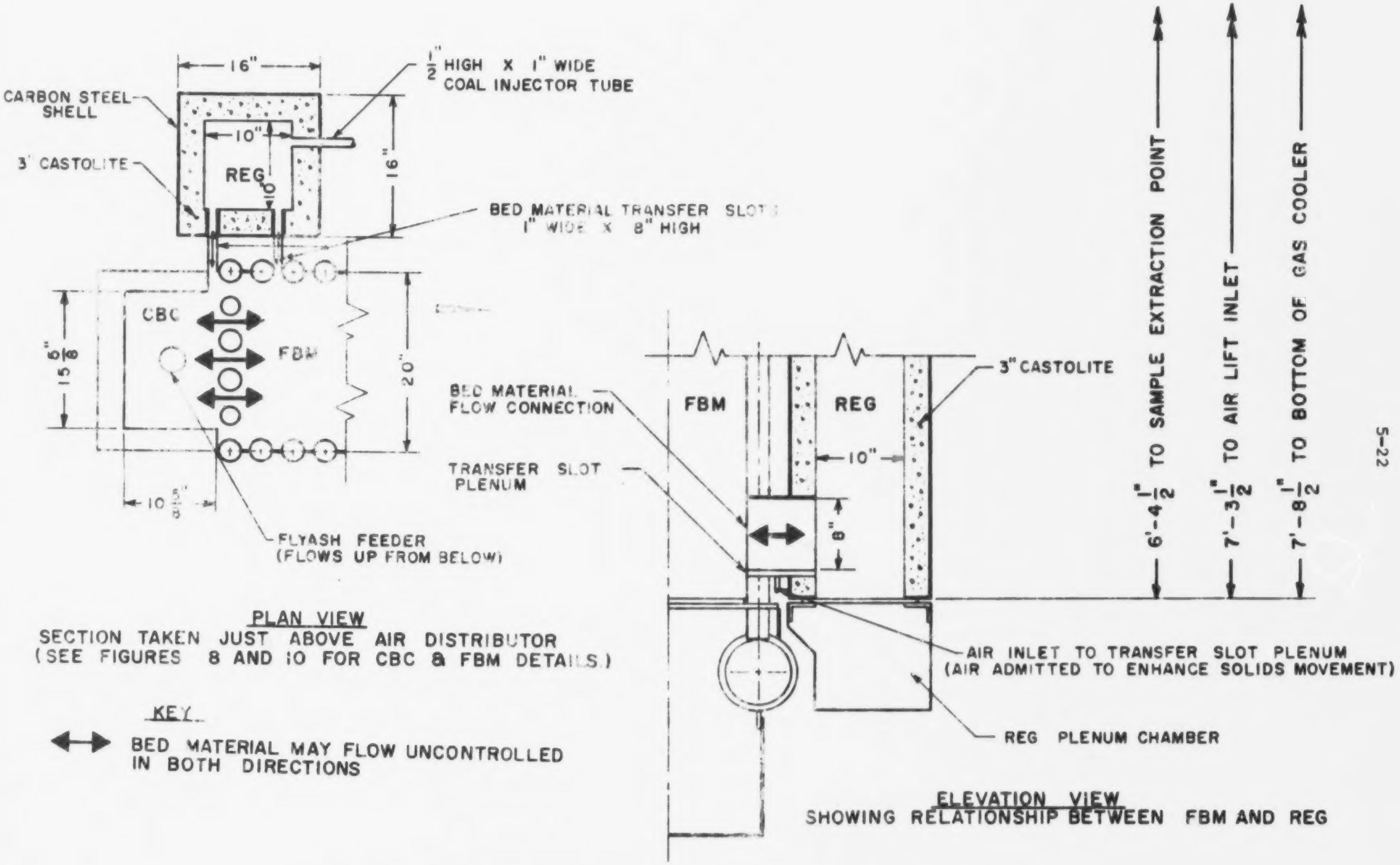
The pilot-scale regenerator, designated the REG, was an appendage to the side of the FBM. It was to operate parallel with the FBM and CBC, at a higher temperature than the FBM, but with a common bed. The design of this system was based on the results of the previous testing with the CBC regeneration operation. The REG cross-section consists of a refractory cement box with a carbon steel shell. It was felt unnecessary to use 446 alloy steel. The REG grid area is 10 x 10 inches, 0.7 ft^2, designed to produce a flue gas with a much higher SO_2 content than the CBC at comparable superficial gas velocity. The grid pattern is identical to the FBM and CBC and at the same elevation. Two bed transfer slots are provided. (See Figure 11A). Bed material removed from the FBM via the gravity bed feed (See Figure 11B) and blown into the REG can return via these bed transfer slots to either the FBM or CBC. Figures 11A, 11B and 11C each show the relationship between the three vessels.

When the air lift is not used bed material moved between the three vessels via gravity.

A bed sample pipe and valve are provided to remove REG bed material for analysis.

The REG coal feeder is a square, horizontal ½-inch x 1-inch tube similar to the FBC operation blowing across the FBM/REG transfer slot. This aspiration is believed to aid bed material transfer, but may have also speeded slag buildup in the REG. Further design, optimization and testing are needed.

A gas sample exit, airlift entrance and tubular flue gas cooler are provided similar to the second CBC flue layout. In order to conserve program funds, no REG dust collector was purchased, but its dust collected in the FBM collector. (See Figure 11B)


16"
CARBON STEEL SHELL
3' CASTOLITE
10"
REG
10"
16"
$\frac{1}{2}$" HIGH X 1" WIDE COAL INJECTOR TUBE
BED MATERIAL TRANSFER SLOTS 1" WIDE X 8" HIGH
CBC
FBM
15 $\frac{5}{8}$"
20"
10 $\frac{5}{8}$"
FLYASH FEEDER (FLOWS UP FROM BELOW)
KEY
BED MATERIAL MAY FLOW UNCONTROLLED IN BOTH DIRECTIONS
FBM
REG
3" CASTOLITE
BED MATERIAL FLOW CONNECTION
10"
TRANSFER SLOT PLENUM
8"
AIR INLET TO TRANSFER SLOT PLENUM (AIR ADMITTED TO ENHANCE SOLIDS MOVEMENT)
REG PLENUM CHAMBER
6'-4$\frac{1}{2}$" TO SAMPLE EXTRACTION POINT
7'-3$\frac{1}{2}$" TO AIR LIFT INLET
7'-8$\frac{1}{2}$" TO BOTTOM OF GAS COOLER

PLAN VIEW
SECTION TAKEN JUST ABOVE AIR DISTRIBUTOR
(SEE FIGURES 8 AND 10 FOR CBC & FBM DETAILS)

ELEVATION VIEW
SHOWING RELATIONSHIP BETWEEN FBM AND REG

FIGURE 11A. PHYSICAL INTERCONNECTIONS BETWEEN FBM, CBC, AND REG.

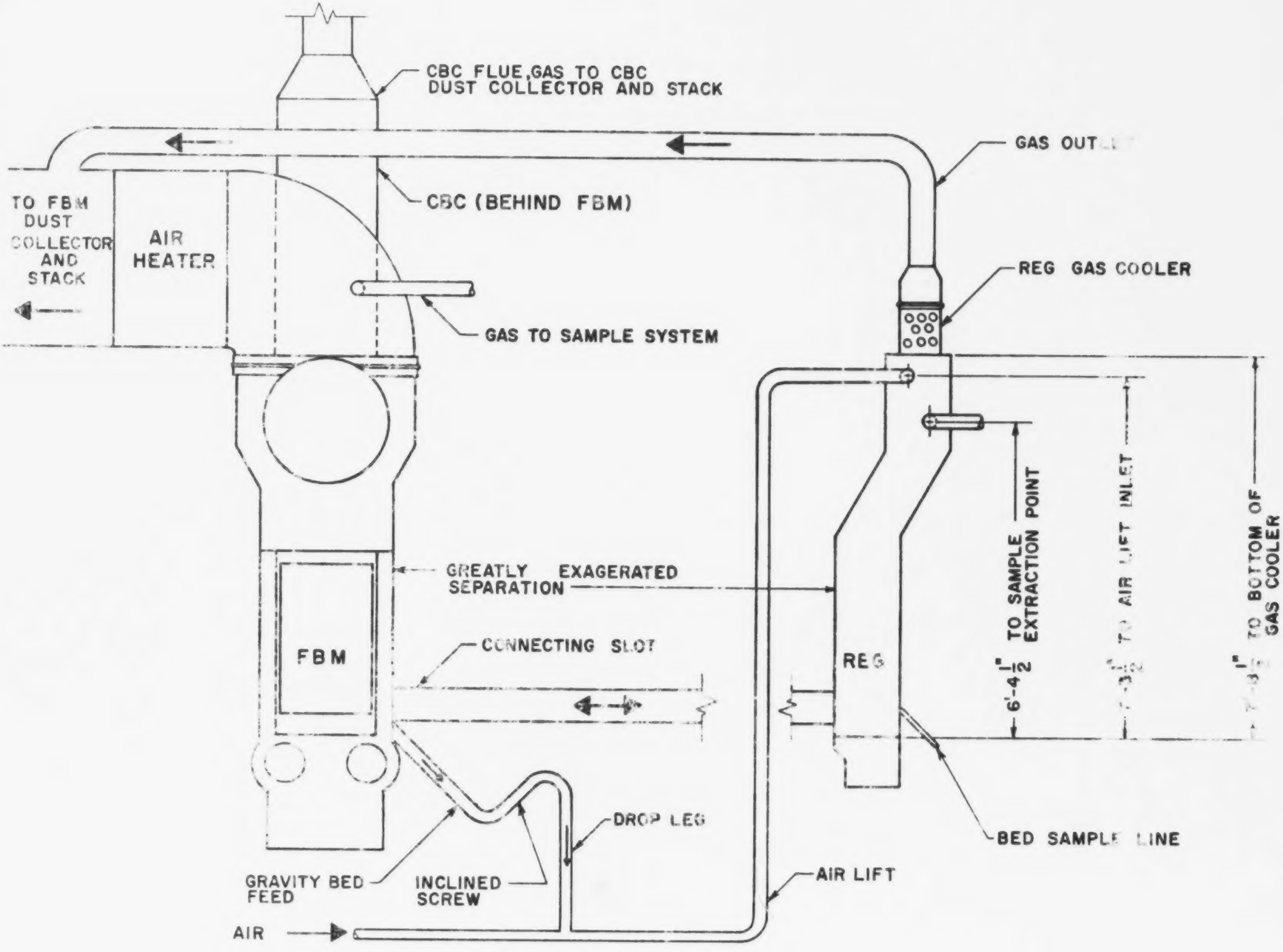

CBC FLUE GAS TO CBC DUST COLLECTOR AND STACK
GAS OUTLET
TO FBM DUST COLLECTOR AND STACK
AIR HEATER
CBC (BEHIND FBM)
REG GAS COOLER
GAS TO SAMPLE SYSTEM
6'-4 1/2" TO SAMPLE EXTRACTION POINT
7'-3 1/2" TO AIR LIFT INLET
TO BOTTOM OF GAS COOLER
GREATLY EXAGERATED SEPARATION
FBM
CONNECTING SLOT
REG
DROP LEG
BED SAMPLE LINE
GRAVITY BED FEED
INCLINED SCREW
AIR LIFT
AIR

FIGURE 11B. FLOW SCHEMATIC FOR FBM AND REG.

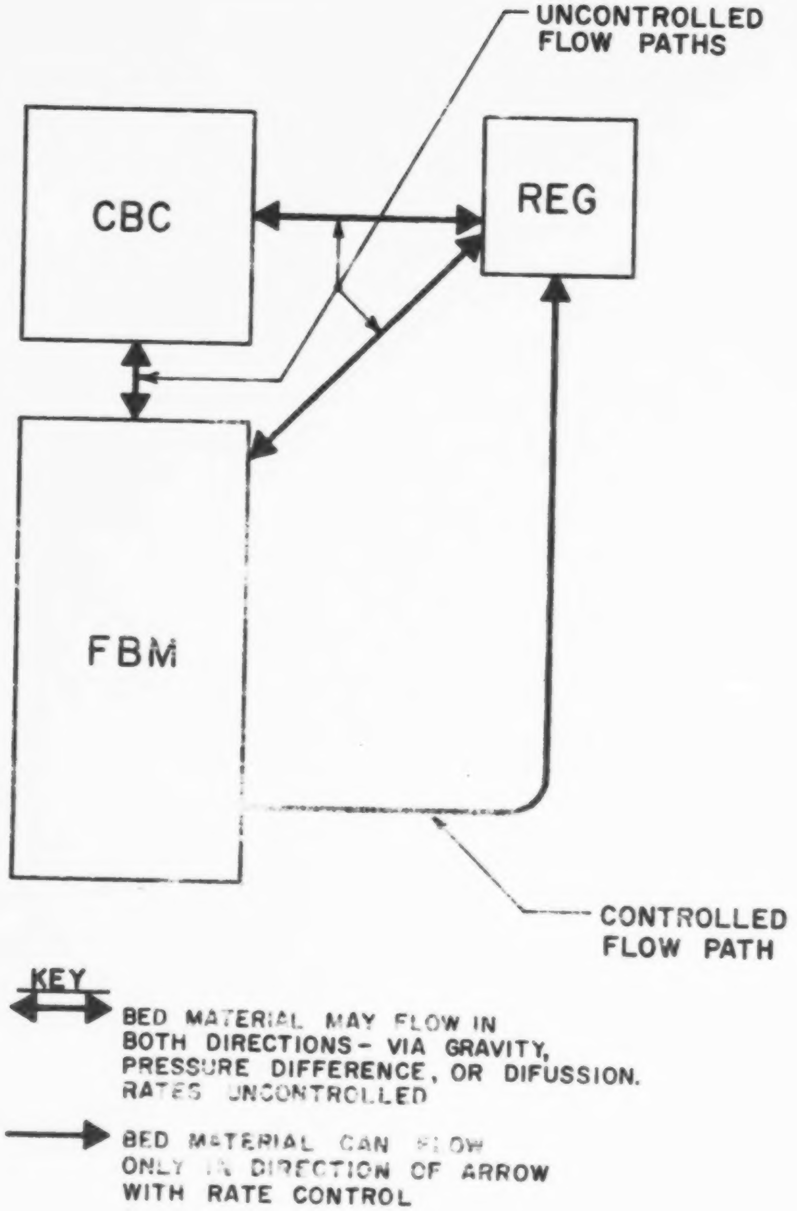

UNCONTROLLED FLOW PATHS
CBC
REG
FBM
CONTROLLED FLOW PATH
KEY
BED MATERIAL MAY FLOW IN BOTH DIRECTIONS - VIA GRAVITY, PRESSURE DIFFERENCE, OR DIFUSSION. RATES UNCONTROLLED
BED MATERIAL CAN FLOW ONLY IN DIRECTION OF ARROW WITH RATE CONTROL

FIG. II-C. FLOW PATHS OF BED MATERIAL FOR FBM, CBC, & REG.

The test procedure involved igniting the FBM and CBC beds in the usual manner and then pulling fluidized bed from the REG (through the sample outlet tube), until coal combustion became stabilized in the REG. The desired velocity, and gas compositions were then created.

In the demonstration runs when the REG was operated, it was fed fine coal (-8 mesh). We recommend that tests to demonstrate an optimal coal size for the REG be performed.

5.5 Instrumentation

Emissions of sulfur dioxide, nitric oxide, carbon dioxide and hydrocarbons were monitored continuously*. Infrared analyzers (Beckman 215) were used to monitor sulfur dioxide, carbon monoxide, and nitric oxide. Carbon dioxide was measured, using a Beckman 7C thermal conductivity analyzer. Hydrocarbons were detected with a flame ionization analyzer (Beckman 109A), using methane as the calibration gas. The signal output of each of these units was displayed on strip chart recorders.

The gas transfer system used with these analyzers is sketched in Figure 12. The system permitted rechecking of calibrations and zero settings on any of the four units at any time during the test by switching from sample gas to reference and zero gases at the rotameter valves. The sample gas was drawn from the hot flue gas stream through a sintered stainless steel filter and conditioned to remove water to a 32°F dew point. The sample gas was again filtered before entry into the analyzers to prevent possible contamination of the optical cells and the hydrogen burner. Wet assay was also used for NO_x (PDS procedure) and SO_x content of process gas samples.

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* Carbon monoxide was also monitored in the extended FBM runs.

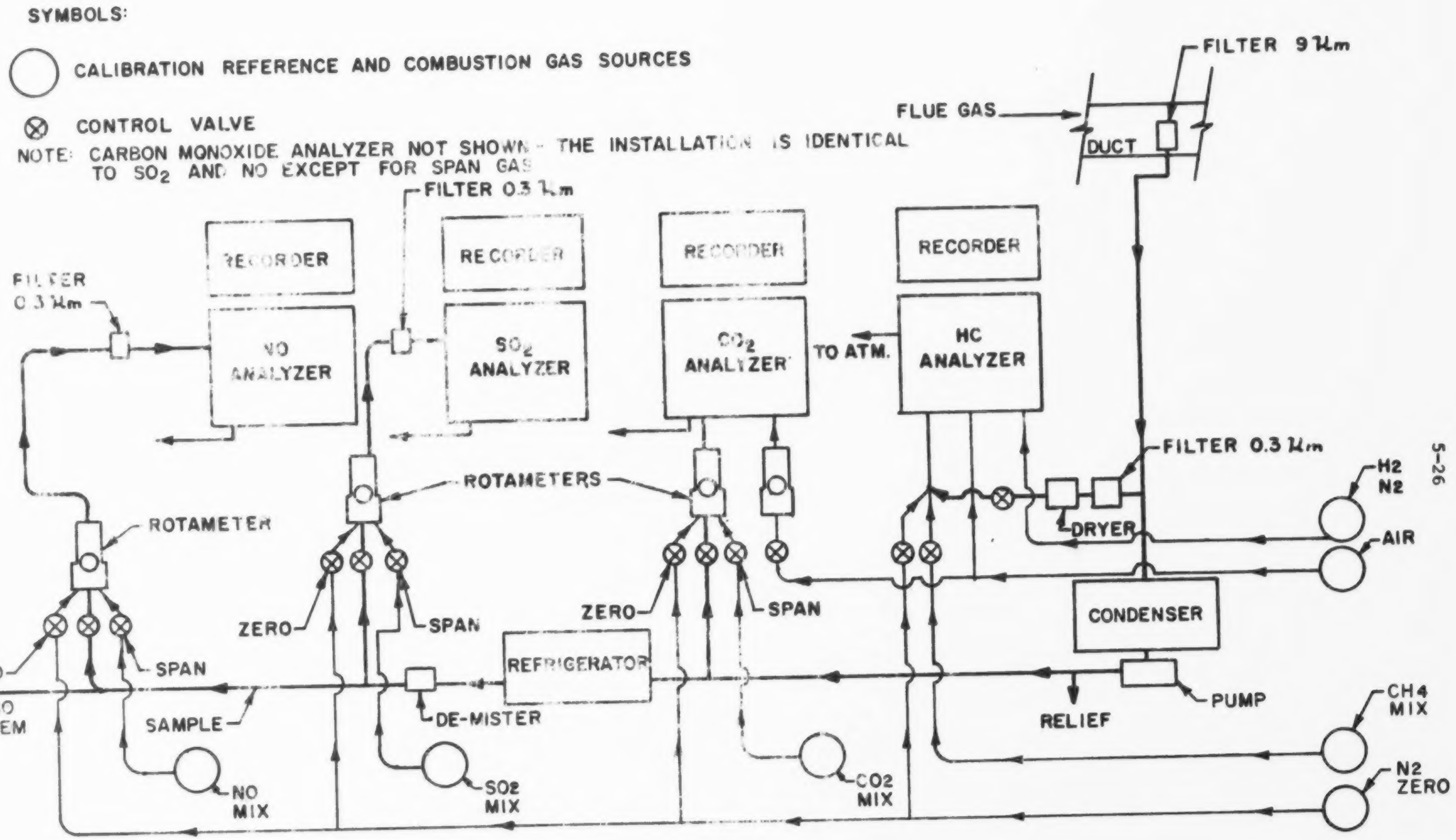
SYMBOLS:
CALIBRATION REFERENCE AND COMBUSTION GAS SOURCES
CONTROL VALVE
NOTE: CARBON MONOXIDE ANALYZER NOT SHOWN - THE INSTALLATION IS IDENTICAL TO SO2 AND NO EXCEPT FOR SPAN GAS
FILTER 9 μm
FLUE GAS
DUCT
FILTER 0.3 μm
RECORDER
RECORDER
RECORDER
RECORDER
FILTER 0.3 μm
NO ANALYZER
SO2 ANALYZER
CO2 ANALYZER
TO ATM.
HC ANALYZER
FILTER 0.3 μm
H2 N2
ROTAMETERS
ROTAMETER
DRYER
AIR
ZERO
SPAN
ZERO
SPAN
CONDENSER
ZERO
SPAN
REFRIGERATOR
PUMP
TO CO SYSTEM
SAMPLE
DE-MISTER
RELIEF
CH4 MIX
NO MIX
SO2 MIX
CO2 MIX
N2 ZERO

FIGURE 12. SCHEMATIC OF GAS TRANSFER SYSTEM FOR CONTINUOUS MONITORING OF SULFUR DIOXIDE, NITRIC OXIDE, CARBON DIOXIDE, AND HYDROCARBONS.

Methods to assay chloride content of flue gases are presently under development. Chloride content with lime beds is expected to be very small*.

The hot FBC gas sample was drawn into the instrument room from the horizontal FBC exhaust duct which extended overhead.

In sampling the FBM flue gas, special precautions were necessary because of the possibility of infiltration of dilution air in the duct above the unit. Also, the poor instrument response which would result from drawing a small sample a long distance (~60 feet) from unit to instrument room was undesirable.

A system was devised to draw a large hot gas sample from the FBM, just above the first gas cooler, pass it through a dust collector, and then through a loop above the instrument room. The sample tube was a [illegible]" pipe with sections screw-fitted and welded. The system was driven with an I.D. fan located at the discharge to atmosphere. A schematic drawing of the system is shown in Figure 13. The CBC and REG gas conduit systems were identical in design. The CBC sample line in the initial CBC configuration was located at the 62" level. For modified CBC, see Figure 10. The REG gas sample line was located below the airlift entrance.

Particulate emissions were monitored with the isokinetic probe and filter system described in Reference 2. The probe design permits equalization of of internal and external static pressures to match the sampling velocity with the stream velocity. Locations of sampling points in the FBC, FBM, and CBC test systems were indicated in Figures 6, 12, and 13, respectively.

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* Cl_2 and HCl in a filtered flue gas stream would be absorbed in aqueous NaOH. The chloride content would be determined colorimetrically with mercuric chloranilate. The absorbance at 530 nm would be correlated with chloride content of calibration samples. With a 15 cu ft. gas sample, 2 ppm chloride sensitivity should be achievable.

Carbon dioxide was continuously monitored, using a self-referencing thermal conductivity analyzer (Beckman Model 7C, range 0-25%). The span calibration gas contains 16% CO_2. The reference and downscale calibration gas is purified air. Instrument cabinet temperature is controlled at 130°F*. Bailey oxygen analyzers (Type OC1530A) were used as operating devices to indicate oxygen concentration in the FBC, FBM, and CBC and REG flue gases. During a test period, the air input rate was held constant and the coal rate adjusted to maintain the oxygen concentration at the desired value. The Bailey instruments have been calibrated periodically with O_2, N_2, and CO_2 mixtures and found to be very reliable. The flue gas oxygen was also verified, using the standard Orsat technique, which determined also carbon dioxide and carbon monoxide**. When the FBM/CBC was operated, a separate oxygen analyzer served each system. When the FBM/CBC/REG was operated, the REG and CBC shared an O_2 analyzer intermittently. Other instruments were shared by the three units by switching from one sample loop to the other***.

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* We found this instrument to be more sensitive to SO_2 than to CO_2, and also to have a negative response to CO.

** Limit of detection 0.1%.

*** This procedure was only partly satisfactory when using the CBC as a regenerator. An SO_2 instrument transient of 30 min was observed when 5000 ppm SO_2 gas (CBC) was replaced with [illegible] ppm (FBM) SO_2 gas. NO, CO, CO_2 and HC responses were immediate, however.

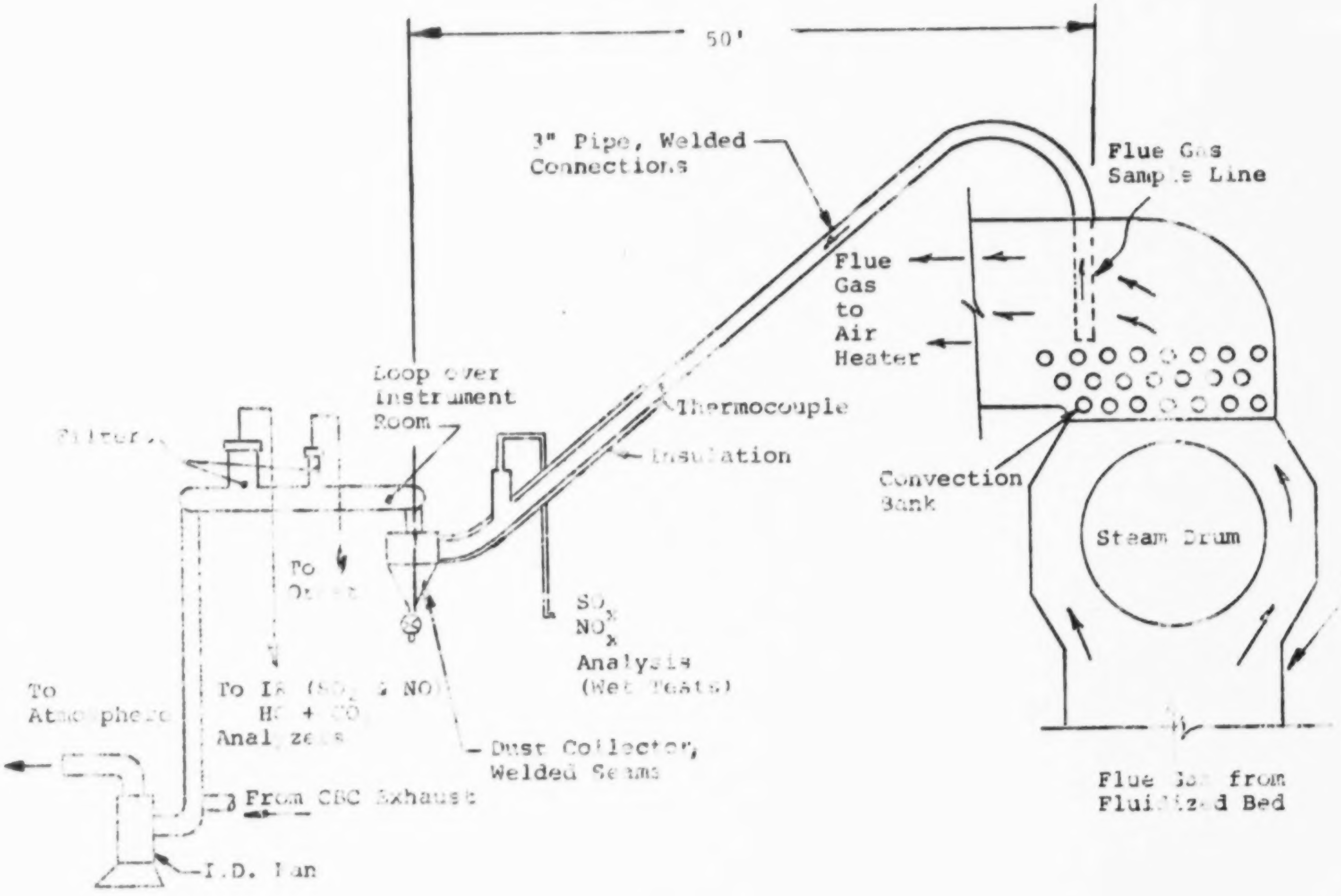
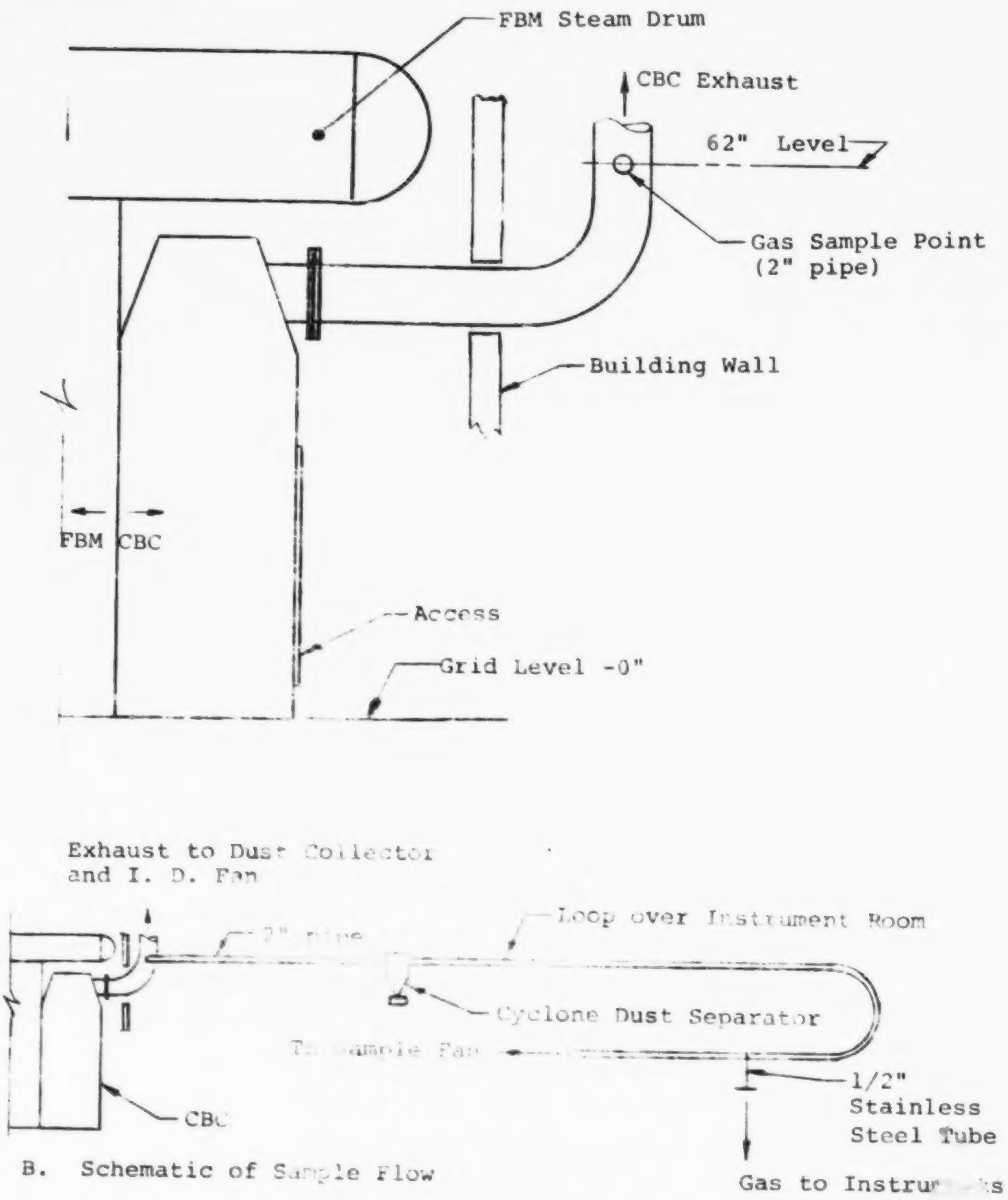

50'
3" Pipe, Welded Connections
Flue Gas Sample Line
Flue Gas to Air Heater
Thermocouple
Insulation
Loop over Instrument Room
Filters
To Orsat
Convection Bank
Steam Drum
FBM
SO_x NO_x Analysis (Wet Tests)
To IR (SO_2 & NO) HC + CO Analyzers
To Atmosphere
Dust Collector, Welded Seams
From CBC Exhaust
I.D. Fan
Flue Gas from Fluidized Bed

FIGURE 13. SCHEMATIC OF THE FBM GAS SAMPLING SYSTEM


FBM Steam Drum
CBC Exhaust
62" Level
Gas Sample Point
(2" pipe)
Building Wall
FBM
CBC
Access
Grid Level -0"
Exhaust to Dust Collector
and I. D. Fan
2" pipe
Loop over Instrument Room
Cyclone Dust Separator
To Sample Fan
1/2"
Stainless
Steel Tube
CBC
Gas to Instruments

B. Schematic of Sample Flow

FIGURE 14. SCHEMATIC OF CBC GAS SAMPLING SYSTEM

Temperatures in the bed and at various other points in the system were recorded remotely on a Honeywell multipoint recorder (cycle time 2 min). A multiple switch panel was used to connect the multipoint recorder input to either the FBC or FBM-CBC systems, as required. Locations of thermocouples in the systems are indicated in Appendices A and C. West dial meters with multipoint temperature selector switches were used as operating indicators, in parallel with the chart recorder. The West meter readings are usually 25°F above the chart record.

The infrared analyzers and the hydrocarbon analyzer were periodically calibrated with gas mixtures supplied by vendors. The concentration of the active components in the calibration gases was checked after delivery to the laboratory. The methane mixture was analyzed by the National Bureau of Standards; a report is shown in Reference 2. This gas, containing 12[illegible] ppm CH_4, was used to calibrate a second methane mixture before it was depleted.

The sulfur dioxide calibration gas was analyzed with a peroxide absorption train. Concentrations of 2650 and 2530 ppm were used in the program. The NO gas was analyzed by PDS technique. Analysis of the nitric oxide calibration gas is also given in Reference 2.

The output signal of the NDIR sulfur dioxide analyzer (0-5000 ppm) varies in a nonlinear manner with SO_2 concentration. [illegible] instrument was checked by precision dilution of the known calibration gas. The curve was found to be correct except for a slight deviation at the low end of the range. The calibration curve and check points are

are given in Reference 2. The combined contribution of H_2O and CO_2 is about 100 to 220 ppm with this analyzer depending on the dew point achieved. The calibration curve was used without correction since the deviation is not more than 1% of full scale. When SO_2 levels exceed 5000 ppm, N_2 dilution gas can be added through a separate flowmeter. When flows of N_2 and sample are equal, a scale deflection of "5000 ppm" can then be interpreted as "1% SO_2". In test 171 H and later tests, the SO_2 analyzer was modified to have three ranges: 0-920 ppm, 0-5000ppm and 0-5%, SO_2.

The calibration curve for the nitric oxide NDIR analyzer is given in Reference 2. The contribution of water vapor to the signal output is significant with this analyzer. The water vapor correction determined by the vendor (180 ppm) was checked by testing a dry gas in the analyzer for comparison with refrigerated ambient air. A correction of 100 ppm was noted and incorporated in the data reduction. The range of this unit is 0-1000 ppm NO. This analyzer was inoperative during Tests C-321, 2, 3, and 171-H.

The carbon monoxide NDIR analyzer is a triple range instrument, zero to 0.35, 1.0 and 2.7%. CO_2 and H_2O give negligible interferences.

Carbon, hydrogen, and sulfur analyses of solid materials were performed using equipment listed in Appendix D. Calcium was analyzed by digestion followed by EDTA titration. Some sulfur analyses (bed material*) were performed gravimetrically by barium chloride precipitation (Eschka).

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* The Leco procedure is time consuming with many of the bed material samples due to porcelain formation.

5.6 Materials

Coal used in FBC tests was Powhatan 7.1% C, 4.5% S, 6.1% ash, 13,123 Btu/lb higher heating value.

Three coals were used in FBM testing; these are the Powhatan containing 4.5% S and 71% C; another also Powhatan, containing 3.3% S and 73.5% C; the third Rivesville coal containing 12.4% ash, 71% C and 3.85% S. FBC and FBM coals were crushed to ¼" top size.

Limestone used was commercially available No. 1359 double screened -8+20 mesh material normally sold as poultry grit.

Salt used was mostly table salt. Its size consist was: on 25 U.S. mesh, 0.1%; on 30, 1.0%; on 50, 85.2%; on 70, 11.9%; on 80, 0.9%; on 120, 0.2%. Some FBC testing was done with road salt having ¼" top size.

6. RESULTS OF BENCH AND PILOT SCALE TESTS

6.1 Reduction of Emissions of SO_2

The prime purpose of the bench [illegible] program has been to evaluate [illegible] of a fluidized-bed combustor, incorporating limestone materials to control SO_2 emission. FBC operation is a batch process in which a charge of limestone is calcined, reacted with SO_2, and finally as an option, can be discarded or can be regenerated by increasing coal rate so that hotter, less oxidizing conditions are created. A regenerated charge of lime could then be run through additional sorption-regeneration cycles if desired. Such batch cyclic operation resembles, but is not congruent with, tandem FBM/CBC sorption-regeneration operation.

During the period covered by this report, sintered ash bed operation with fine limestone addition has been de-emphasized as a SO_2 control means, due to the poor limestone utilization experience at several installations using this technique.

The FBC and FBM units contain provisions for fly ash reinjection; no reinjection studies were conducted during the period covered by this report. Some minor improvement in limestone utilization might have been achieved via reinjection since the fly ash contained some unreacted lime.

6.2 FBC Tests

Of the extensive FBC test series originally planned, only four were conducted. Available funds were then transferred to the FBM long duration testing program.

FBC testing was conducted with -8+20 mesh, 1359 limestone beds and ¼" x 0 Powhatan 4.5% sulfur coal, with and without activity enhancement additives. (See Appendix E for background information.) Run No. C-321 was initiated with low coal and air rates and a low bed level to expedite calcining. After establishment of coal combustion and calcining, further bed material was added. Following completion of most of the calcining, as evidenced by the flue gas CO_2 level strip chart record, data acquisition began. (See Table 1.) Bed temperature was 1580°F. Since the FBC hood insulating liner used in previous fly ash combustion tests was still in place, heat transfer was limited and high excess air was needed to limit bed temperature. Beginning at 200 ppm flue gas SO_2 built up gradually to 2100 ppm* after 5.5 hours of combustion (including calcining time). Analyses of bed material samples show a corresponding buildup of bed sulfur content (See Figure 15)**. At the relatively low ingredients rates and bed level, apparent combustion efficiency is 90% or more, although fly ash collection efficiency appears low, especially in Conditions 1-4. A sulfur balance (see Appendix F) was compiled for this run showing the amounts accumulated in fly ash, flue gas, and bed material. The principal sulfur inventory at short run times (<7 hours) in the

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* If no acceptor were present, the coal and air rates correspond to a maximum of 2750 ppm SO_2.

** Bed calcining period was about 2 hours. More precise definition is not possible since a CO_2 emission tail-off of about 30 min. usually follows stone feed cutoff.

TABLE 1. FBC TEST DATA SUMMARY

Test No. C-321	-1	-2	-3	-4	-5	-6	-7
1. Time (hours)	2	2½	3.8	N/A	5.5	6.6	7
2. Air rate, lb/hr/ft^2	710	710	710	710	710	710	710
3. Bed temperature, $^{\circ}$F	1580	1610	1580	1560	1570	1560	1540
4. Bed depth, in. (1359 Limestone)	16	16	16	16	16	16	16
5. Bed particle size -8+20							
6. Coal input rate lb/hr/ft^2	51	51	48	48	48	48	48
7. Carbon input, lb/hr	31.2	31.2	29.1	29.1	29.1	29.1	29.1
8. Fly ash output, lb/hr	N/A	2.8	2.2	2.5	4.2	5.1	7.5
9. Output C content, %wt.	27.6	47.8	48.8	46.	53.9	56.7	40.
10. Output S content,% Average 2.26	2.71	2.52	2.53	1.88	2.08	1.95	2.15
11. Output Ca content,%	20.5	12.0	6.1	5.1	5.0	5.3	9.76
11a. Ratio Ca/S in output	7	4.7	2.4	2.7	2.4	2.7	4.5
12. Carbon output, lb/hr	N/A	1.33	1.07	1.15	2.26	2.9	3.
13. Carbon burned	N/A	29.9	28.	28.	28.6	26.2	26.1
14. Combustion eff.,%	N/A	96	96.3	96	92	90	90
15. Superficial velocity, ft/sec	9.7	9.8	9.7	9.6	9.6	9.6	9.5
16. Cooling probe position, 4 probes 100% inserted throughout the test							
17. Fuel heat input, KBtu/hr	575	575	535	535	535	535	535
18. Sulfur input, lb/hr	1.98	1.98	1.85	1.85	1.85	1.85	1.85
19. NaCl rate, lb/hr	0	0	0	0	0	4.0	4.0
20. Sulfur output, lb/hr	.076	.070	.056	.047	.087	.099	.161
21. $CaSO_4$/CaO ratio in fly ash, wt.	0.48	0.86	2.6	2.0	6.	2.0	0.93
22. Bed sulfur content, %wt.	2.58	N/A	5.86	N/A	7.57	8.96	9.75
22a. Bed Ca content, %wt.	51.7	N/A	51.8	N/A	48.7	49.6	40.6
22b. Limestone feed rate	0	0	0	0	0	0	0
Flue Gas Composition:							
23. CO_2, %vol.	10.6	10.9	11.1	10.7	11.1	11.	10.9
24. O_2, %	6.8	6.4	6.8	7.	6.7	7.	6.0
25. CO, %	0	0	0	0	0	0	0
26 SO_2, ppm	200	450	900	1500	2100	500	650
27. HC, ppm	3	15	0	0	0	0	0

TABLE 1. (continued)

Heat Balance, KBtu/hr	-1	-2	-3	-4	-5	-6	-7
28. Wall loss	13.0	10.0	13.4	11.2	12.6	10.7	10.7
29. Hood loss (insulated)	28.0	26.7	35.7	30.0	33.6	28.5	28.5
30. Probe loss	209	209	201	194	176	179	164
31. Flue gas loss (at T bed)	263	269	261	257	257	255	251
32. Carbon loss	11	18.7	15.1	16.2	32.	41.	42.
33. Ash heat content	0.5	0.5	0.4	0.4	0.7	0.8	1.2
34. Total loss	524	534	527	509	512	515	498

Notes: N/A = not available. "Output" material is that collected by the dust collector.

Bed weight post test not available. 16 in. of used bed usually weighs 64 lb.

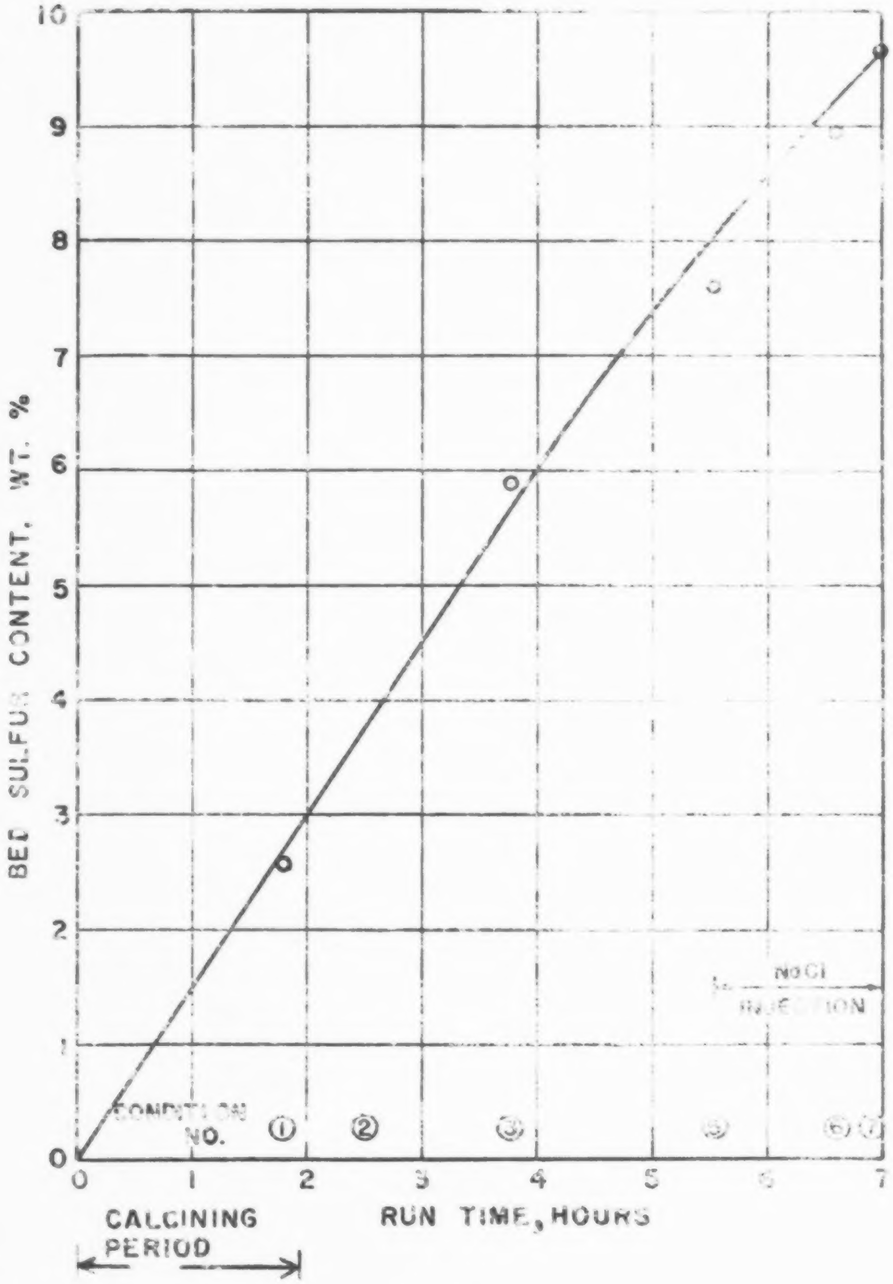

FIGURE 15 BED SULFUR ACCUMULATION VS TIME
FBC RUN — C-321

absence of either regeneration or the enhancement additive is laid down in bed material (sulfated lime). The variable flue gas SO_2 levels (often 1000-2000 ppm) during the heavy calcining period account for most of the deviation between input (fed in coal) and total outputs, since the percent recovery increases with time as the flue gas SO_2 increases.

Starting at 5.7 hours, coarse salt, ¼" top size, addition was initiated, with a screw feeder dropping salt into the coal feed air stream. Salt was ¼" road de-icing type material. The pressure drop across the feed screw, plus backflow of fly ash, caused intermittent loss of NaCl flow. Nevertheless, SO_2 levels with the aged lime bed are reduced dramatically by salt addition. At the bed temperature and air rate used, the calculated NaCl vapor rate in flue gas is about 1 lb/hr, based on published vapor pressure data, and the flue gas isokinetic sampler displayed a high salt content. Probably the chloride content of salt is tied up as calcium chloride in bed material or fly ash, and sodium is tied up as sodium calcium sulfate in bed material and fly ash.

The apparent effects of the salt addition to an aged, unregenerated bed containing CaO and 32% $CaSO_4$ are:

a) The SO_2 emission rate from coal combustion is cut back to rate approximately that with freshly calcined CaO. (See Table 1, Condition 6.)

b) The rate of fly ash collection is increased. It is hypothesized that the fly ash contains a dust of $CaNa_2(SO_4)$ or similar double salt whose crystal lattice parameters are sufficiently different from calcined limestone that particle outer shell strength is reduced and surface removal occurs naturally by attrition in the fluidized bed, leaving exposed active calcium oxide.

c) The rate of sulfur laydown in the bed material is restored to a value approximating that with freshly calcined CaO (See Figure 15); i.e.,CaO activity has been enhanced.

d) The rate of darkening of fresh bed material (the brown discoloration that may be due to iron pickup as $CaFe_2O_4$) is arrested.

Several names have been proposed for this process:

1) SO_2 Acceptor Process with Chemical Scrubbing
2) SO_2 Acceptor Process with Chemical Attrition

Following completion of Run C-321, the FBC hood liner insulation was removed and a vibrator added to the salt feeder to improve feeder reliability. Run C-322 was then initiated with a fresh bed to test limestone bed operation at higher coal rate, lower flue gas O_2, and deeper bed conditions. Run temperature was 1600°F. Flue gas SO_2 built up at about the same rate as previously. (See Table 2) At the higher coal and air rates, fly ash amounts collected were considerably greater and combustion efficiencies somewhat lower. The significantly higher hydrocarbon levels in this test (120 ppm) are attributed to both lower O_2 and reduced freeboard height, due to the deeper bed in a constant height apparatus.

Coarse salt addition was attempted following growth of SO_2 to 1000 ppm*, a lower level than in the previous test. Due to the higher low-bed pressure in this test, fly ash blowback occurred rapidly and only intermittent

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* If no acceptor were present, the coal and air rates correspond to a maximum of 3800 ppm SO_2.

TABLE 2. FBC TEST DATA SUMMARY

Test No. C-322	-1	-2	-3	-4	-5	-6	Avg
1. Time (hours)	1.9	2.9	3.9	4.9	5.9	6.9	
2. Air rate, lb/hr/ft^2	800	780	760	770	770	780	
3. Bed temperature, °F	1590	1630	1595	1620	1610	1620	
4. Bed depth, in. (Limestone)	22	N/A	N/A	N/A	N/A	N/A	
5. Bed particle size, -8+20	N/A	N/A	N/A	N/A	N/A	N/A	
6. Coal input, lb/hr/ft^2	74	74	65	81	77	76	
7. Carbon input, lb/hr	45	45	39.8	49.6	46.9	46.5	
8. Fly ash output, lb/hr	20,7	20.7	25.5	22.8	31.7	24.9	
9. Output C content,%	27.6	22.4	*29.4	26.6	32.4	29.7	
10. Output S content,%	2.26	2.52	2.42	2.57	2.18	2.23	2.3
11. Output Ca content,%	17.0	18.5	15.7	16.3	19.2	19.7	
11a. Ratio Ca/S in fly ash	7.5	7.4	6.5	6.4	8.8	8.8	
12. Carbon output, lb/hr	5.7	4.6	*7.5	6.1	10.2	7.4	
13. Carbon burned	39.3	40.4	32.3	43.5	36.7	39.1	
14. Combustion eff., of carbon,%	87.3	90	N/A	88	78.3	84.1	85.
15. Superficial velocity, ft/sec	10.8	10.6	10.4	10.5	10.5	10.6	
16. Cooling probe position	2 in	3 in	3 in	4 in	4 in	4 in	
17. Fuel heat input, KBtu/hr	830	830	731	915	862	855	
18. Sulfur input, lb/hr	2.86	2.86	2.52	3.15	2.97	2.95	
19. Sulfur output	0.47	0.52	0.61	0.59	0.69	0.56	
20. Bed sulfur content, wt.%	2.36	4.39	6.26	7.10	8.70	9.71	
20a. Bed calcium content	49.1	51.3	46.5	47.4	46.2	[illegible]	
20b. Ratio Ca/S in bed	21	11.7	7.4	6.7	5.3	4.5	
20c. Salt feed rate	0	0	0	0	0	0	

Flue Gas Composition:

	-1	-2	-3	-4	-5	-6
21. CO$_2$, %vol.	14.8	14.8	15.	15.	14.7	14.6
22. O$_2$, %	3	3.1	3.2	2.9	3.	3.2
23. CO, %	0	0	0	0	0	0
24. SO$_2$, ppm	200	1000	1800	**1800	2750	2900
25. HC,ppm	150	120	180	180	170	120

TABLE 2. (Continued)

[illegible]	[illegible]	-2	[illegible]	[illegible]	[illegible]	[illegible]
26. Wall loss	2	3	2	3	3	[illegible]
27. Hood loss (uninsulated)	181	191	153	176	166	[illegible]
28. Probe loss	183	208	169	187	197	[illegible]
29. Flue gas loss (at T bed)	306.2	308.7	285.2	310.2	298.7	[illegible]
30. Carbon loss	80.4	65	106	86	144	[illegible]
31. Hydrocarbon loss	2.6	2	2.9	3	2.8	2
32. Ash heat content	6.4	6.4	7.7	7	9.7	[illegible]
33. Total loss	762	784	725	772	821	[illegible]

* FBC banked 10 min. between -2 and -3.

** 7.6 lbs makup bed added between -3 and -4 (Raw basis)
Initial bed mass 167.9 lbs (raw); 105 lbs (calcined)
Final bed mass 88 lbs after test (as result of elutriation, [illegible]

Note: No salt addition (see text). "Output" material is that collected by the dust collector feed of small amounts of salt could be achieved, with only small perturbing effects on SO_2. The FBC unit was then banked, the salt hopper cleaned out, and a fresh batch of fine salt introduced. The unit was restarted and fine salt addition attempted. Although no fly ash blowback occurred, the 1/80 HP variable speed motor was unable to operate with this much frictional drag. and either stalled or stripped its set screws. Salt feed was therefore abandoned, and the balance of the test conducted as a straight limestone bed test.

Since there was evidence of some bed material being carried over and collected with the fly ash, at the high combustion rate and bed level conditions, a small amount of fresh limestone bed material (7.6 lb) was fed very slowly over a 40 min. period between Conditions 3 and 4. The effect on flue gas CO_2, due to calcining at this low makeup rate was minimal. In a full scale boiler using this once-through type of process, withdrawals of [illegible] would be made in the form of [illegible] material as well as fly [illegible] limestone would be required.

The bed sulfur content and sulfur balance [illegible] length of the calcining period as [illegible] CO_2 level strip chart record.

Run No. C-323 was initiated with low coal and air rates, and a low bed level to expedite calcining. After establishment of coal combustion and calcining, further bed material was added. Following completion of most of the calcining, as evidenced by the flue gas CO_2 level strip chart record, data acquisition began (See Table 3). Bed depth was 14 in. (static). Due to the limited heat

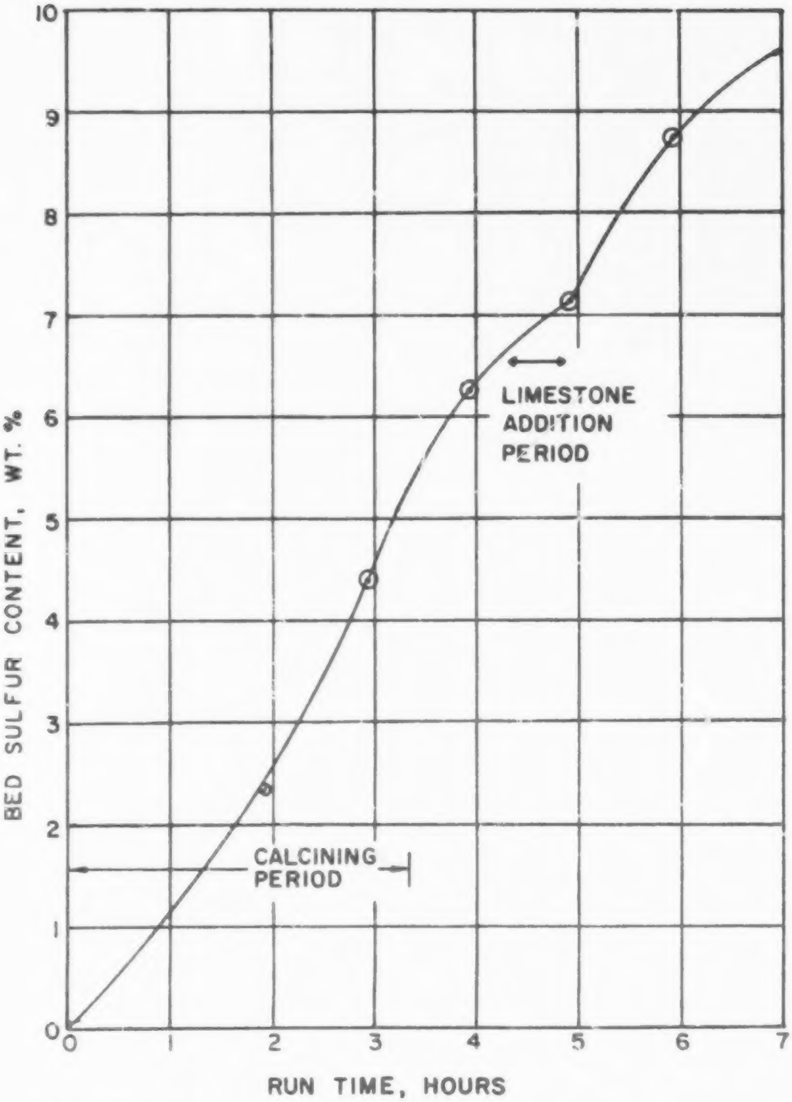

FIGURE 16. BED SULFUR CONTENT VS. TIME
FBC RUN C-322 (LIMESTONE BED)

TABLE 3. FBC TEST DATA SUMMARY

Test No. C-323	-1	-2	-3	-4	-5	Avg
1. Air rate, lb/hr/ft^2	790	800	790	800	800	
2. Bed temperature, ^{o}F	1650	1600	1610	1530	1525	
3. Bed depth, in.	14	14	14	14	14	
4. Bed particle size, -8+20						
5. Coal input, lb/hr/ft^2	58	58	55	79	79	
6. Input carbon content, %wt.	71					
7. Carbon input, lb/hr	35.5	35.5	33.4	48.3	48.3	
8. Fly ash output, lb/hr	2.9	6.	4.6	7.7	7.7	
9. Output C content, %wt.	43.2	56.0	46.2	51.8	47.	
10. Output S content, %wt.	2.46	2.08	2.05	2.21	2.43	2.25
11. Output Ca content, %wt.	13.2	7.24	15.13	7.62	10.85	
12. Ratio Ca/S	5.3	3.5	7.3	3.4	4.4	
13. Carbon output, lb/hr	1.2	3.3	2.1	4.0	3.6	
14. Carbon burned	34.3	32.2	31.3	44.3	44.7	
15. Combustion eff.,%	-	90.7	94.	92.	92.5	
16. Superficial velocity, ft/sec	11.3	11.0	11.1	10.7	10.7	
17. Cooling probes position	4 in	4 in	4 in	3 in	3 in	
18. Fuel heat input, KBtu/hr	662	662	662	900	900	
19. Sulfur input, lb/hr	2.25	2.25	2.11	3.06	3.06	
20. NaCl rate, gal/hr	0	0	0	0	1.4**	
21. Sulfur output, lb/hr	.071	.125	.094	.17	.18	
22. Bed sulfur content, %wt.	3.98	5.70	6.85	8.61	8.26*	
23. Bed Ca content, %wt.	53.6	49.4	49.2	41.6	37.3	
23a. CaCO$_3$ Feed Rate	0	0	0	0	0	
Flue Gas Composition:						
24. CO$_2$, %vol.	13	11.8	12.	12.5	12.3	
25. O$_2$, %	4.8	6.	5.8	4.8	5.	
26. CO, %	0	0	0	0	0	
27. SO$_2$, ppm	600	950	1850	660	660	
28. HC, ppm	10	30	0	0	0	

* Aged sample, hydrated

**26 wt% aqueous solution

TABLE 3 (Continued)

Heat Balance, KBtu/hr:	-1	-2	-3	-4	-5
29. Wall loss	8.3	5.9	6.5	3.2	N/A
30. Hood loss	111	100	110	119	N/A
31. Probe loss	214	201	191	143	N/A
32. Flue gas loss at 1400°F	256.4	259.1	254.3	272.1	N/A
33. Carbon loss	17	47	30	57	N/A
34. Ash heat content	0.9	1.8	1.4	2.3	N/A
35. H_2O ΔH	0	0	0	21	N/A
36. Total loss	608	616	594	618	N/A

transfer at this low bed level, temperature could not be maintained below 1610°F (above the optimum value for SO_2 absorption), at O_2 levels of about 5.8%. Beginning at 580 ppm, flue gas SO_2 built up gradually to 1850 ppm* after 3.7 hours of combustion (including calcining time). Analyses of bed material samples show a corresponding buildup of bed sulfur content (See Figure 17). At the moderate ingredients rates and low bed level, apparent combustion efficiency is 91% or more, although fly ash collection efficiency appears low, especially in conditions 1 and 3. A sulfur balance (See Appendix Figure F-3) was compiled for this run, showing amounts accumulated in fly ash, flue gas, and bed material. As in previous non-regeneration tests, the principal sulfur inventory at short run times (<6 hours), in the absence of the enhancement additive, is laid down in bed material (sulfated lime).

Starting at 4 hours, salt solution addition was initiated. Pressurized saturated salt solution (35 grams NaCl in 100 ml H_2O) was fed from a pressurized vessel through a valve and flowmeter and dribbled into the coal-feed air stream, at the coal introduction point. This method is less malfunction-prone, and smoother than the screw feeding of coarse or fine salt, but imposes a heat loss on the system in the form of water vaporization. Visual inspection of the coal-feed air mixed stream during operation disclosed no buildup of evaporated salt deposits in the coal-feed system. As previously, the SO_2 levels in the flue gas with the aged lime bed are reduced dramatically by salt addition. The additional

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* If no acceptor were present, the coal and air rates correspond to a maximum of 3680 ppm SO_2.

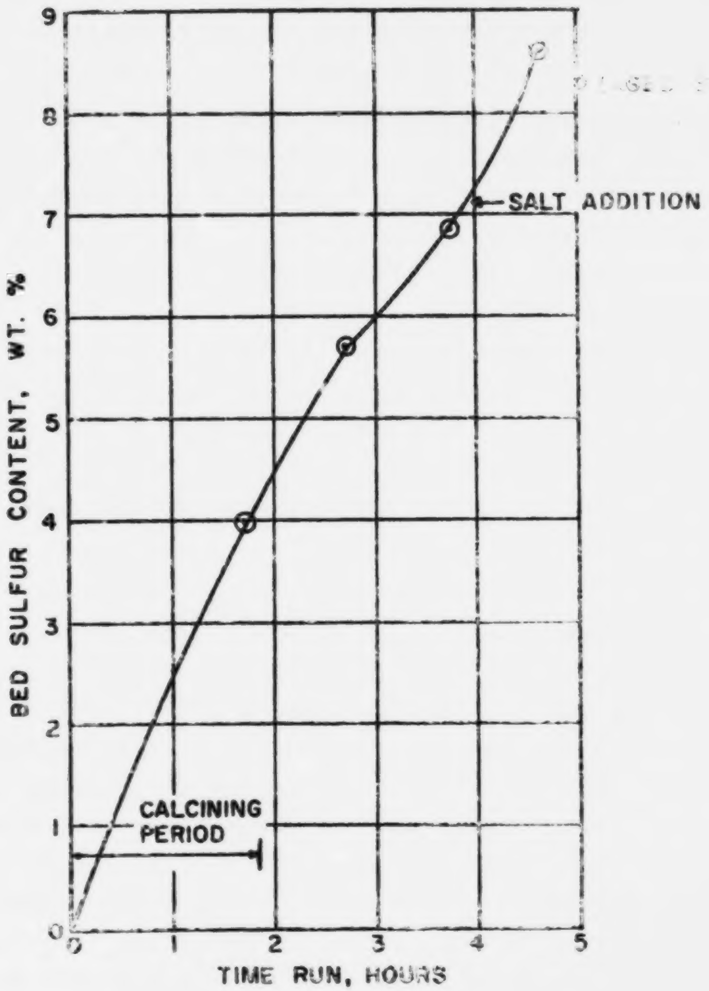

FIGURE 17. BED SULFUR CONTENT VS. TIME FBC RUN C-323

effects of salt noted previously are also prominent -- increased fly ash production and increased rate of sulfur laydown in bed material. As already stated on p. 6-5, vaporized salt condenses in the cooled flue gas as an aerosol.

The test was abruptly terminated prematurely by a city-wide power failure. The condition 5 bed sample was inadvertently aged about 1 month before analysis, leading to hydration and carbonation, and hence low indicated S and Ca contents, but the ratio Ca/S follows the orderly progression of the other conditions. Compared to Test No. C-322, the hood heat loss is lower in this test due to the lower bed level and resulting in increased freeboard height with less bed particle impingement on the cooled surfaces -- simultaneously, there is a lower carbon loss and higher combustion efficiency, with reduced flue gas hydrocarbon levels.

Following completion of Run C-323, the FBC solid salt feeder was reinstalled in the system with a new low friction feed screw. Run C-324 was then initiated with a fresh bed to test limestone bed operation at lower air rates, lower flue gas O_2, and deeper bed conditions. Run temperature was 1540°F. Beginning at 150 ppm, flue gas SO_2 built up gradually to 650 ppm* after 4.4 hours of running (See Table 4). Combustion efficiencies averaged about 89.5%. Fine salt addition (screw fed into the coal feed air stream) was initiated following the condition No. 3, at an SO_2 level lower than in the previous test. Ever at the low air rate (low feed air manifold pressure) salt feed was erratic. Several periods of down time

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* If no acceptor were present, the coal and air rates correspond to a maximum of 3100 ppm SO_2.

TABLE 4. FBC TEST DATA SUMMARY

Test No. C-324	-1	-2	-3	-4	-5	-6	-7	-8	Avg.
1. Air rate, lb/hr/ft^2	720	730	720	740	700	710	720	710	
2. Bed temperature, °F	1590	1560	1550	1530	1470	1520	1520	1540	
3. Bed depth, in.	17	17	17	17	17	17	17	17	
4. Bed particle size, -3+20	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	
5. Coal input, lb/hr/ft^2	53	58	60	58	53	59	59	47	
6. Input carbon content, %wt.	71	N/A	N/A	N/A	N/A	N/A	N/A	N/A	
7. Carbon input, lb/hr	32.6	35.5	36.7	35.5	32.6	35.8	35.8	28.4	
8. Fly ash output, lb/hr	N/A	8.3	8.8	8.	16.2	16.6	8.5	20****	
9. Output C content, %wt.	39.3	39.8	45.5	50.6	35.3	44.3	26.4	33.7	
10. Output S content, %wt.	2.31	2.60	2.54	2.42	2.89	3.89	3.49	3.35	2.47***** 3.4******
11. Output Ca content, %wt.	22.9	16.22	11.80	15.18	18.95	12.70	22.50	22.40	
12. Ratio Ca/S in output	9.9	6.2	4.7	6.2	6.5	3.3	6.4	6.6	
13. Carbon output, lb/hr	3.3	3.3	4.0	4.1	5.7	7.3**	2.3	6.7****	
14. Carbon burned	29.3	32.2	32.7	31.4	26.9	28.5	33.5	21.7	
15. Combustion efficiency, %	90	91	89	88.5	83	N/A	94	77****	89.3
16. Superficial velocity, ft/sec	10.4	10.3	10.1	10.6	9.5	9.9	9.0	10.0	
17. Cooling probes position	3 in	4 in	N/A	N/A	N/A	N/A	N/A	N/A	
18. Fuel heat input, KBtu/hr	609	662	688	662	609	666	666	530	
19. Sulfur input, lb/hr	2.07	2.25	2.34	2.25	2.07	2.27	2.27	1.8	
20. NaCl rate, lb/hr	0	0	0	3*	3	2	2	0	
21. Sulfur output, lb/hr	0.19	0.22	0.22	0.19	0.47	0.65	0.30	0.67	
22. Bed sulfur content, %wt.	3.39	4.85	6.45	7.75	9.86	7.76	11.0	8.42	
23. Bed Ca content, %wt.	53.2	56.8	54.7	51.4	46.1	45.8	45.4	41.4	
24. Ratio Ca/S in bed	15.7	11.7	8.5	6.6	4.7	5.9	N/A	N/A	

See Footnotes on page following.

TABLE 4. (Continued)

Test No. C-324	-1	-2	-3	-4	-5	-6	-7	-8	Average
Flue Gas Composition:									
25. CO_2, Vol. %	13	12.9	13.3	13.2	13.	13.5	13.	14.4	
26. O_2, %	4.4	4.	3.7	2.8	3.	3.4	4.8	2.2	
27. CO, %	0	N/A	N/A	N/A	N/A	N/A	N/A	N/A	
28. NO, ppm ***	390	300	250	270	180	240	200	140	
29. SO_2, ppm	280	430	560	430	430	460	490	1300	
30. HC, ppm	5	40	90	2	10	2	2	170	
Heat Balance, KBtu/hr									
31. Wall loss	2	3.1	2	2	2	2	2	N/A	
32. Hood loss	124	115	116	127	93	108	108	N/A	
33. Probe loss	170	195	189	195	139	171	184	N/A	
34. Flue gas loss at 1400°F	234	239	237	243	226	231	237	227	
35. Carbon loss	47	47	57	58	81	103	33	95	
36. Ash heat content	2.5	2.5	2.6	2.3	4.5	4.8	2.5	5.8	
37. Total loss	580	602	604	628	546	620	567	~622	

NOTES:

* Salt feed erratic.

** Down time: 5 min. 2:25 to 2:30; 3.5 min. 2:48 to 2:52.

*** NO recorder erratic.

**** 18 minute condition.

***** Without salt.

****** With salt.

Small sorbent addition made intermittently; see text.

occurred in attempts to feed salt uniformly. With salt addition, the SO_2 level was reduced to a fairly uniform 430 ppm. Simultaneously there was a dramatic reduction in flue gas hydrocarbon content, to essentially zero ppm, attributable to the catalytic effect of salt on vapor phase combustion. Fly ash rate is increased by salt addition, as previously. Note that the chemical attrition theory implies that 58 grams of salt can yield as much as 278 grams of calcium sodium sulfate ($CaNa_2(SO_4)_2$). Some salt is lost at 1500°F as vapor. The observed ratio, 8 lb additional fly ash to 3 lb salt fed, is consistent with the theory. There was a temporary nitric oxide (flue gas) reduction at the start of salt feeding, but no lasting effect. In later FBM tests with salt addition, salt reduced the NO output.

At six points during the test, limestone bed material increments were fed over time intervals of 1, 2, or 3 minutes, and at the rate of 1 lb/min, to reduce bed temperature and make up bed level after attrition losses. In each pulse, typical bed temperature drop was 50°F; CO_2 increases 2.5 to 3 percent, both effects due to calcining. In addition, hydrocarbon readings abruptly increase by 200 to 300 ppm; O_2 level temporarily drops by about 2.5 percent in conformity with the hydrocarbon rise. NO levels drop about 100 ppm in accord with the O_2 change; SO_2 temporarily increases 50 ppm in accord with the O_2 shortage, but quickly recovers and lines out at about 30 ppm below the value preceding limestone pulsing. The source of hydrocarbons during limestone addition may be the result of an organic impurity in the raw stone, an effect on the coal feed entering via the same port, or an effect on bed hydrodynamics.

Fly ash sulfur contents in this run average 2.5% without salt, and 3.4% when under the influence of salt. Bed sulfur content rises smoothly to 9.86% at 6 hours of running; beyond this time the sulfur values become erratic, probably due to aging of samples and hydration prior to analysis. The apparent ratios Ca/S in conditions 5, 7, and 8 in the fly ash are higher than those in the bed material; this is an interesting effect, and was noted in conditions 5 and 6 of Run C-322 where no salt had been fed; it implies that the bed particle core is enriched in sulfur and that sulfur (as SO_2, SO_3, $SO_3^=$,$SO_4^=$ or other form) under coal combustion conditions is mobile in the particles and that the surface, which erodes, becomes depleted in sulfur as it is enriched in iron, and possibly other contaminants. This mobility has been named D.E.S., or Dynamic Exchange of Sulfur. Experiments required to underatand D.E.S. were described in an unpublished progress report (11). Electron microprobe studies at Argonne National Laboratory (12) appear to confirm the existence of D.E.S.

Sulfur balance data for this test are shown in Table F-1. The sulfur balances in conditions 1 through 5 are good; those in conditions 6 through 8 are unsatisfactory, in line with the erratic bed sulfur content values mentioned previously. As previous, most sulfur is retained in the bed.

Following condition 7, salt was cut off and excess air reduced to drive up SO_2 emission. This was a short condition, and SO_2 rose by 800 ppm to 1300 ppm, at which point the test was terminated. Hydrocarbons rose to 170 ppm.

A discussion of the bed particle size vs time data recorded in these tests is presented in Section 6.6.

Following completion of test No. [illegible], the program financial and manpower resources were devoted solely to larger scale FBM testing (see next section).

6.3 FBM/CBC Tests

As discussed in Section 4, Introduction, the demonstration goals of the FBM testing were:

a) 90% or better SO_2 removal from the primary cell flue gas;

b) 3 to 4% (vol.) or more SO_2 in regeneration section flue gas;

c) 98% or better carbon burnup efficiency in the overall system.

6.3.1 Test with the Short CBC/Regenerator

The integrated FBM/CBC boiler system was operated and data recorded in a preliminary SO_2 Acceptor Process test. A crushed limestone bed was used for SO_2 capture in the FBM. Simultaneous fly ash and coal feed to the CBC were used. During the final portions of the test, the CBC stack gas contained between 1.5 and 2.4 percent SO_2, the concentration depending inversely on O_2 level, indicating regeneration of circulating bed material at about 1870°F. The concentration factor, defined as the ratio of the SO_2 concentration in the regenerator region off-gas to that in the sorbent region off-gas, was on the order of 20. SO_2 output from the FBM was above 900 ppm.

The purpose of Test No. B-18 was to determine the physical limitations of the current FBM/CBC system when operating in the SO_2 Acceptor Process MK I (two cell) mode. In this mode, one region (the primary cell) of a fluidized-bed boiler is operated under conditions favorable to sulfur acceptance by lime ($T_{BED} < 1550^{\circ}F$, Excess $O_2 > 2\%$) while a smaller, adjacent region is operated under conditions favorable to sulfur rejection ($T_{BED} > 1800^{\circ}F$, Excess $O_2 < 4\%$). This high temperature zone would also burn the fly ash from the primary zone. Bed particles are made to circulate between the two regions, for example, by means of diffusion* or under the influence of a pressure differential. This test was made with E. Ohio, Pittsburgh No. 8 seam 'Powhatan" coal with 12.1% ash and 4.5% sulfur contents. A crushed (-8+20) 1359 limestone bed was used. Run duration was about ten hours. The intent of the test was to capture sulfur and generate a partially sulfated lime bed in the boiler, and then, demonstrate simultaneous regeneration of aged bed material in a hotter, less oxidizing zone (the CBC acting as a regenerator). To expedite circulation of sulfated boiler lime to the CBC and regenerated lime back to the boiler, additional intercommunicating slots were cut in the baffle which separates the two regions. A total of six square inches was provided compared to the 4 inches used in Ref. 1. Due to the increased intercommunication area, the heat sink effect of the FBM made it necessary during the test to drive up the CBC temperature by feeding coal to supplement the fuel value of the carbon bearing fly ash. Additional air required to burn the coal brought the CBC air rate to 1190 lb/hr, close to 20% of the total for the system. This contrasts with the 10% design value. A low amount of air to the regeneration zone is very important in the regenerative mode of operation, since a concentrated SO_2 stream is desired.

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* In portions of later tests, diffusional circulation was supplemented by mechanical circulation.

The combination of the high volatile coal being used in the FBM and the less erosive nature of lime bed particles compared to ash bed material appeared to lead to somewhat thicker than normal carbon deposits on the 400°F boiler tubes. This reduces the heat absorption from the bed and raises the bed temperature*. Normally this is counteracted by raising the bed level until the desired equilibrium is restored. Attempts were made to provide adequate boiler waterwall heat transfer by increasing bed mass. However, due to the inadequate freeboard and high air velocity in the CBC (short configuration), much loss of bed material was experienced, with associated plugging of the CBC gas sample line. The overbed baffle screen, which was effective in the short CBC with an ash bed, did not seem as effective in knocking down regenerated lime particles.

Test data are summarized in Table 5, corresponding to about the ninth hour of running.

Due to the coal being fed at various rates to the CBC, and to the large fraction of lime particles in the cyclone product, the heat balance around the unit is incomplete and carbon-burnup data are not presented.

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* This problem did not recur in subsequent tests.

TABLE 5. FBM-CBC TEST SUMMARY

Test No. B-18	A	B
FBM Heat Balance		
Fuel heat input, KBtu/hr	5886	N/A
Boiler steam gain	3248	3400
Circulating H_2O absorption	1295	N/A
Flue gas loss (includes CBC)	911	N/A
Thermal efficiency,%	77.2	N/A
FBM		
Bed material	calcined limestone	
Air rate, lb/hr	4774	N/A
Superficial velocity, ft/sec	7.6	N/A
Coal rate, lb/hr	450	N/A
Bed temperature, °F	1610	1630
Fly ash output, lb/hr	125	N/A
Fly ash carbon content, %	56.7	N/A
Fly ash sulfur content, %	2.57	N/A
Fly ash calcium content	7.15	N/A
Carbon combustion eff.,%	78.	N/A
Flue Gas Concentration		
O_2, vol. %	5.	4.
SO_2 ppm	970	1040
NO ppm	360	380
CBC		
Air rate, lb/hr	1102	N/A
Superficial velocity, ft/sec	14.5	N/A
Coal feed	Yes	Yes
Bed temperature, °F	1850	1880
Flue Gas Concentration		
O_2, vol. %	1.3	~0.5
SO_2, %	1.55	2.44
NO ppm	330	260

Due to inability to maintain the desired high bed level, the bed was usually above 1550°F, the temperature at which better SO_2 capture would occur (see Test Log, Appendix H).

During most of the test, the hydrocarbons analyzer gave erratic output and results are not reported. The failure of the hydrocarbon analyzer is not related to the operation in the SO_2 acceptor mode, and did not recur in subsequent tests.

As shown in Appendix H, flue gas SO_2 levels of up to 1.5 to 2.4 percent were achieved from the CBC/regenerator. At the CBC air rate used, the 2.4 percent value exceeds the steady state value (sulfur input to the combined system); i.e., the total bed inventory was experiencing a depletion in sulfur and the 2.4 percent level would not have been maintained indefinitely. Bed sulfur content data are shown in Figure 18.

Based on the results of this test, a new CBC/regenerator as discussed in Section 5.3, was designed having sufficient freeboard to eliminate regenerated lime particle entrainment, as well as sufficient refractory insulation, giving improved heat economy to avoid the necessity of coal feed to maintain the desired temperature. It appears likely that the bed intercommunication area provided in this test (6 in^2) was excessive, inhibiting achievement of the desired temperature differential. Two slots, each 2 in^2, are probably sufficient; if at different levels, with bed materials of different sulfur content and temperature a thermosiphon effect will occur. Manometer data indicated expanded bed heights of 16" (FBM) and 22" (CBC). Since the overbed pressure differential was 2.9" H_2O, there was probably some admixture of high SO_2 CBC gas with the FBM flue gas, partially accounting for the 970-1040 ppm SO_2 readings off the FBM.

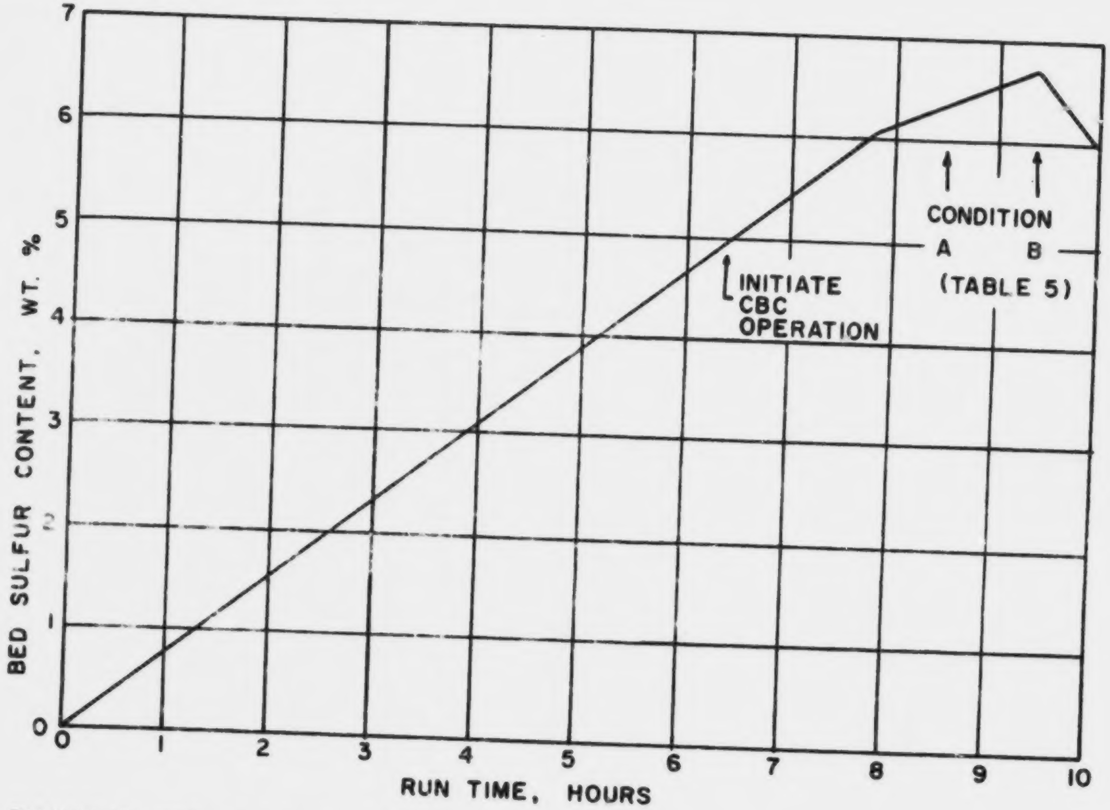

FIGURE 18. BED SULFUR CONTENT VS. TIME, FBM RUN B-18 ILLUSTRATING REGENERATION

The principal reaction occurring in the regenerator process is:

$$C + CaSO_4 + \frac{1}{2}O_2 \rightarrow CO_2 + CaO + SO_2$$

which is endothermic by 750 cal/gram of sulfur. The sulfur release is about 9 lb/hr. This endotherm is more than offset by the carbon-burnup exotherm (7800 cal/gram).

A number of factors can lead to an energy deficit within the fluidized-bed for operation in the 2000°F range. Energy can be lost through the walls, in heating bed particles which enter the system at a lower temperature, in heating inert dust and the carbon which does not burn within the bed, and in heating nitrogen and excess oxygen. An energy balance in a high temperature bed can be achieved by preheating the reactants, by heat exchange against products and/or by minimizing energy losses.

6.3.2 FBM/CBC Extended Run (Run 168H): Test with the Tall CBC/Regenerator*

6.3.2.1 Test Operation and Burnup Results

The purpose of this test with the new CBC (See Figure 8) was to continuously regenerate lime bed material and demonstrate sulfur capture in the FBM of 90% or better based on coal input, and to study CBC carbon burnup and simultaneous flue gas SO_2 output as a function of CBC conditions (temperature, oxygen level, additive coal feed, carbon content of input fly ash). The test duration was 72 hours plus initial time (about 8 hours) establishing burnup/regeneration conditions. Test data are summarized in Table 6. Coal was the "Powhatan" 71% C, 4.5%S, 12.1% ash material. The 90% capture goal was easily met and exceeded. Fifty-nine capture measurements were logged,

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* This test was preceded by a shakedown run of 28 hours' duration. Coal feed problems were experienced and no data are reported.

TABLE 6. FBM/CBC SYSTEM LOG - 168H RUN

Day	FBM Time	T, °F	Bed Level	Coal Rate lb/hr	Air Rate lb/hr	Fly Ash Carbon, %	Flue Gas SO2, ppm	Flue Gas lb/hr	S Release Rate, lb/hr	% S Capture	Bed S %	Fly Ash* % S	Fly Ash* % Ca	O_2 %	CO %	HC ppm	NO ppm	CO_2 %	Steam lb/hr	CBC O_2 %	CBC Ash C, %	CBC SO_2 %	CBC Air Rate lb/hr
3 8	1808	1550	Low	500	5910	N/A	350	N/A	2.28	89.9	N/A	N/A	N/A	3.8	0.8	550	260	11.9	3600	6.8	N/A	N/A	N/A
	1900	1550	13.	500	6090	46.5	150	N/A	0.9	96.	N/A	1.50	N/A	4.	0.7	3-0	220	11.1	3500	7.8	N/A	N/A	577
	2015	1540	N/A	466	6060	46.6	250	[illegible]	1.75	91.6	7.84	1.71	N/A	4.7	N/A	400	290	11.5	3500	3.5	12.9	1.0	N/A
	2110	1570	N/A	522	N/A	N/A	230	N/A	1.53	93.4	7.84	N/A	N/A	3.5	N/A	800	260	12.5	3400	2.5	N/A	2.4	N/A
	2200	1555	16.6	522	5995	54.5	N/A	6338	1.61	93.1	7.84	1.78	N/A	2.8	N/A	900	260	13.	3550	2.0	39.	2.4	580
	2300	1572	N/A	514	N/A	N/A	220	N/A	1.46	93.6	7.84	N/A	N/A	2.6	N/A	1000	260	13.	3700	1.6	N/A	3.4	N/A
9	0007	1550	N/A	N/A	N/A	55.	N/A	6329	1.54	93.1	7.84	N/A	N/A	3.	N/A	N/A	N/A	N/A	3750	1.1	31.3	0.72	777
NR	0137	1490	15.5	N/A	6408	N/A	N/A	N/A	N/A	N/A	7.84	N/A	N/A	4.5	N/A	N/A	N/A	N/A	3750	0.	N/A	0.4	N/A
	0200	0147	16.	586	N/A	N/A	N/A	N/A	N/A	N/A	[illegible]	N/A	N/A	4.8	N/A	100	250	N/A	3800	0.	N/A	N/A	N/A
	0255	1500	N/A	565	N/A	N/A	N/A	N/A	N/A	N/A	[illegible]	1.89	N/A	3.8	N/A	1000	190	N/A	3700	0.	N/A	N/A	N/A
	0400	1540	15.2	656	N/A	N/A	150	N/A	1.09	95.4	N/A	N/A	N/A	3.	N/A	1000	210	15.	4100	0.	N/A	6.	N/A
NR	0430	1540	15.2	557	6545	40.3	250	N/A	1.79	92.8	3.50	1.78	N/A	3.5	N/A	500	210	11.8	4100	5.	N/A	0.5	914
	0610	1490	16.	557	6588	52.7	300	6959	2.3	90.0	3.68	1.70	N/A	2.5	1.8	500	290	11.	4100	2.5	30.3	N/A	912
	0720	1470	N/A	N/A	N/A	53.9	N/A	6954	N/A	N/A	4.95	1.74	N/A	3.4	N/A	350	N/A	11.	3800	2.	26.2	2.4	N/A
	0747	1450	N/A	N/A	N/A	N/A	300	N/A	1.46	93.7	N/A	N/A	N/A	3.5	N/A	450	270	10.5	3800	2.6	N/A	1.6	N/A
	0840	1450	17.	N/A	6494	58.6	150	N/A	1.08	95.1	5.60	N/A	N/A	3.2	1.2	95	260	10.8	3800	1.8	N/A	1.5	929
	0949	1490	17.	525	6160	55.5	200	6500	1.43	94.3	5.20	N/A	5.30	2.7	1.5	240	260	10.7	3900	3.	24.	1.0	949
	1051	1480	17.	500	6365	N/A	250	N/A	1.75	92.2	6.88	N/A	N/A	2.8	1.1	250	260	10.9	3900	2.7	18.5	1.5	920
NR	1152	1475	17.	560	6895	52.5	310	7291	2.30	90.1	8.15	N/A	6.63	2.7	1.2	240	280	11.	4100	2.4	17.6	0.4	867
	1312	1480	18.	417	6538	33.5	380	6880	2.885	84.6	Salt in	N/A	5.42	3.0	0.9	280	260	10.7	4000	2.6	15.8	1.25	730
NR	1404	1485	18.	N/A	N/A	57.4	200	6862	1.52	93.8	N/A	N/A	N/A	3.	2.8	400	200	11.	3900	4.5	7.75	>0.5	~900
	1524	N/A	N/A	N/A	Banked	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	1644	1500	17.	545	6892	55.	~350	7237	~2.79	~88.6	N/A	N/A	N/A	3.4	0.85	200	280	11.	3800	0.	38.7	4.	617
	1745	1505	17.	N/A	7152	43.6	300	7536	2.49	89.5	N/A	N/A	N/A	4.5	0.8	180	290	11.5	3850	0.	35.8	4.	897
	1845	1495	17.	N/A	7142	57.5	250	7488	2.06	91.6	N/A	N/A	N/A	5.7	0.44	80	350	11.3	3600	1.2	38.1	4.	924
	1945	1550	16.	N/A	7346	N/A	200	N/A	1.63	93.3	N/A	N/A	N/A	5.4	0.31	55	370	11.5	4000	1.2	N/A	3.3	924
	2100	1500	16.	548	N/A	N/A	200	N/A	1.63	93.1	N/A	N/A	N/A	6.7	0.3	20	280	8.	4000	1.6	N/A	0.75	N/A
	2130	1540	16.	548	7207	51.9	230	7618	1.93	92.1	N/A	N/A	8.28	6.4	N/A	30	270	8.2	4000	1.3	10.4	1.6	622
	2300	1590	17.5	548	7365	51.5	250	7732	2.13	91.4	2.82	N/A	N/A	3.2	2.2	1000	260	15.	4000	1.2	26.0	3.	640
	2400	1510	N/A	545	N/A	N/A	280	N/A	2.28	90.7	N/A	N/A	N/A	3.5	N/A	300	320	13.8	4000	1.3	N/A	2.8	N/A
10	0100	1520	17.1	600	7250	N/A	230	N/A	1.85	93.1	N/A	N/A	N/A	3.	N/A	500	310	13.8	4000	1.5	N/A	4.4	609
	0200	1520	N/A	610	N/A	N/A	200	N/A	1.60	94.1	N/A	N/A	N/A	2.6	N/A	300	370	N/A	4000	1.5	N/A	4.1	N/A
	0300	1500	N/A	N/A	Banked	52.8	N/A	N/A	N/A	N/A	N/A	1.84	N/A	N/A	N/A	50	N/A	13.	N/A	2.0	N/A	N/A	N/A
	0400	1480	17.5	N/A	7310	N/A	150	N/A	1.22	95.4	N/A	N/A	N/A	4.	N/A	300	310	12.	3900	2.4	N/A	0.96	850
	0500	1510	16.6	N/A	7562	47.8	170	7973	1.49	94.5	1.40	1.88	N/A	4.	N/A	16-	320	11.6	4200	2.	46.6	0.84	605
	0635	1530	17.3	600	7411	56.	250	7800	2.15	92	1.41	1.68	N/A	3.7	0.102	400	320	13.5	4200	3.2	44.6	1.2	750
	0730	1530	N/A	592	N/A	41.2	260	~7833	2.24	~91.6	N/A	1.54	N/A	3.8	N/A	400	320	13.8	4200	3.	47.7	1.1	N/A
	0821	1510	N/A	N/A	7130	54.6	250	7516	2.07	~92.2	1.24	1.65	N/A	3.6	0.105	350	310	13.6	4100	3.	62.5	1.2	727
	0930	1525	17.5	639	7115	35.4	250	7587	2.09	92.7	1.26	1.76	N/A	4.1	0.17	700	310	13.7	4200	3.5	66.	1.05	727
	1029	1520	17.5	591	7025	57.	240	N/A	1.36	93.0	1.16	1.45	N/A	4.5	0.14	440	320	13.3	4100	N/A	56.5	1.5	727
	1128	1525	18.	576	6930	59.7	260	N/A	1.39	92.3	1.11	1.45	N/A	4.2	0.15	400	320	13.4	4050	N/A	59.6	1.35	722

*Anlysis performed on material collected by dust collectors only.

NOTE: NR = nonregenerating portion of test
N/A = Not Available

TABLE 6. (Continued) Page 2.

Day	Time	FBM S Input in Coal lb/hr	CBC Temp. °F	CBC NO ppm	CBC (CO$_2$ & SO$_2$) %	System C Burnup % Overall	CBC CO %	CBC HC ppm	FBM Limestone Add. Rate lb/hr	CBC Flue Gas lb/hr	CBC Flue Gas Sulfur lb/hr	CBC Coal Rate lb/hr	CBC Bed % S	CBC Fly Ash % S	CBC Fly Ash % Ca	Concentration SO$_2$ Factor CBC/FBM
8	1823	22.5	[illegible]	N/A	N/A	N/A	N/A	N/A	14	N/A	N/A	0	N/A	N/A	N/A	N/A
	1908	[illegible]	2000	N/A	N/A	N/A	N/A	N/A	40	N/A	N/A	0	N/A	N/A	N/A	N/A
	2015	21.	2000	N/A	19.4	97.7	N/A	100	28	~768	8.5	0	N/A	1.10	16.22	40
	2110	N/A	1980	N/A	22.	N/A	N/A	300	85	N/A	16.2	0	N/A	1.41	18.21	104
	2218	23.5	1955	N/A	22.5	92.3	N/A	350	85	612	16.2	0	N/A	(2.45)	N/A	104
	2315	N/A	1950	N/A	23.	N/A	N/A	400	71	N/A	30.7	0	N/A	N/A	N/A	154
9	0001	[illegible]	1840	N/A	19.7	96.7	N/A	300	70	820	6.5	0	N/A	1.89	N/A	33
	0137	N/A	1920	N/A	N/A	N/A	N/A	N/A	73	N/A	N/A	16	N/A	N/A	N/A	18
	0200	N/A	1900	N/A	25.	N/A	N/A	N/A	84	N/A	N/A	16	N/A	N/A	N/A	N/A
	0300	N/A	1890	N/A	25.	N/A	N/A	N/A	91	N/A	N/A	16	4.75	2.04	N/A	N/A
	0400	N/A	1910	N/A	25.	N/A	N/A	100	91	N/A	63.	16	N/A	N/A	N/A	400
	0450	N/A	1900	N/A	17.	N/A	N/A	0	80	N/A	N/A	16	3.18	1.84	N/A	20
	0610	25.	1910	N/A	N/A	~93.3	N/A	N/A	77	953	N/A	16	6.50	1.88	N/A	N/A
	0720	~25.0	1840	N/A	23.	~94.5	N/A	20	77	962	25.4	16	5.31	1.76	N/A	80
	0800	N/A	1920	N/A	23.	N/A	N/A	10	77	N/A	17.	16	N/A	N/A	N/A	80
	0840	N/A	1940	790	22.5	N/A	0.7	15	84	N/A	19.7	16	N/A	N/A	N/A	100
	0949	[illegible]	2000	860	17.5	97.3	1.4	16	0.	1192	13.1	0	5.44	N/A	17.1	50
	1051	[illegible]	2000	1080	20.5	N/A	0.28	16	0 to 22	N/A	19.7	0	6.62	N/A	19.63	67
	1152	[illegible]	[illegible]	1000	17.3	97.5	0.07	26	17	922	3.7	0	6.84	N/A	19.55	131
	1312	[illegible]	1995	1200	21.	97.1	0.07	28	27	745	10.3	N/A	N/A	N/A	N/A	33
	1409	N/A	[illegible]	N/A	N/A	98.7	N/A	N/A	22	976	3.6	0	7.91	N/A	17.95	25
	1524	Banked	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	1644	[illegible]	1930	450	>25.	93.6	0.55	180	36	653	28.8	0	N/A	N/A	N/A	114
	[illegible]	N/A	1920	490	>25.	91.4	0.46	90	60	969	42.7	~66	5.01	N/A	N/A	133
	[illegible]	N/A	1980	500	>25.	90.5	2.2	30	N/A	992	43.7	~27	N/A	N/A	N/A	160
	[illegible]	N/A	[illegible]	[illegible]	23.3	N/A	2.7	22	0	N/A	36.	N/A	N/A	N/A	N/A	165
	[illegible]	N/A	[illegible]	[illegible]	21.	N/A	[illegible]	30	0	N/A	N/A	N/A	N/A	N/A	N/A	38
	[illegible]	[illegible]	[illegible]	N/A	[illegible]	98.2	N/A	50	0	828	14.6	0	N/A	N/A	23.4	70
	[illegible]	[illegible]	[illegible]	[illegible]	25.	95.6	0.4	120	N/A	683	22.6	0	2.47	N/A	N/A	120
	[illegible]	N/A	[illegible]	N/A	25.	N/A	N/A	130	N/A	N/A	21.1	15.8	N/A	N/A	N/A	100
	[illegible]	[illegible]	[illegible]	290	25	N/A	1.7	70	42	N/A	31.5	15.8	N/A	N/A	N/A	191
	[illegible]	[illegible]	[illegible]	N/A	23	N/A	N/A	80	N/A	N/A	29.4	15.8	N/A	N/A	N/A	205
	[illegible]	N/A	[illegible]	N/A	[illegible]	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	[illegible]	N/A	[illegible]	N/A	21.5	N/A	N/A	50	N/A	N/A	9.5	N/A	N/A	N/A	N/A	64
	[illegible]	[illegible]	[illegible]	N/A	[illegible]	[illegible]	N/A	70	N/A	631	6.	29.	0.80	3.19	N/A	50
	[illegible]	[illegible]		N/A	[illegible]	98	N/A	110	N/A	795	10.5	23.3	1.48	2.18	N/A	48
	[illegible]	[illegible]		N/A	[illegible]	[illegible]	N/A	120	N/A	~774	9.4	23.3	N/A	1.90	11.39	43
	[illegible]	N/A	[illegible]	30		~89	0.71	100	N/A	~774	10.3	25.	0.65	2.36	N/A	48
	[illegible]	[illegible]	[illegible]	70	[illegible]	[illegible]	0.90	130	N/A	773	9.0	33.4	0.48	0.84	N/A	42
	[illegible]	[illegible]	[illegible]	10	[illegible]	N/A	0.63	80	N/A	N/A	12.7	33.4	N/A	2.05	N/A	63
	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	N/A	0.84	58	3.0	N/A	11.4	33.4	N/A	1.91	N/A	52

TABLE 6. (Continued)

Day	Time	FBM T, °F	Bed Level	Coal Rate lb/hr	Air Rate lb/hr	Fly Ash Carbon, %	Flue Gas SO2, ppm	Flue Gas lb/hr	S Release Rate, lb/hr	% S Capture	FBM Bed S %	FBM Fly Ash S%	O2 %	CO %	HC ppm	NO ppm	CO2 %	Steam lb/hr	CBC O2 %	CBC Ash C, %	CBC SO2 %	CBC Air Rate lb/hr
10	1243	1535	18.	588	7010	52.0	350	N/A	2.7	89.6	N/A	N/A	3.8	N/A 0.17	N/A 450	N/A 310	13.6	4050	N/A	63.3	1.25	722
	1342	1530	18.	568	6840	50	250	7247	2.0	92.1	1.19	1.57	3.4	0.14	240	280	11.2	4100	1.8	22.6	1.3	805
	1441	1530	18.	558	6960	40.1	260	7361	2.14	91.5	1.45	1.34	3.5	0.18	150	280	11.	4050	1.5	15.1	1.1	827
	1542	1510	19.	N/A	7040	N/A	200	N/A	1.56	93.7	1.88	N/A	5.	0.10	N/A	280	10.	3900	1.8	N/A	0.75	787
	1646	1535	18.	N/A	6740	41.6	310	7097	2.42	89.3	3.98	1.29	4.5	0.11	425	290	10.6	3900	0.5	19.6	1.45	707
	1758	1490	18.	510	6916	47.3	260	7288	2.10	90.8	3.56	1.45	4.2	0.14	550	270	11.	3950	0.1	13.1	2.4	1090
	1857	1480	17.	540	7010	39.2	250	7426	2.05	91.6	3.03	1.46	4.1	0.105	380	290	10.7	4000	0.4	11.9	2.35	937
	1953	1520	17.	540	7066	N/A	225	N/A	1.76	92.7	2.37	1.31	4.6	0.09	300	300	10.6	4050	0.8	N/A	2.0	939
	2045	1470	17.	N/A	7160	N/A	N/A	N/A	N/A	N/A	N/A	N/A	4.5	N/A	250	280	11.3	4000	0.3	N/A	N/A	887
	2130	1470	17.2	N/A	7324	N/A	100	N/A	0.61	96.6	N/A	N/A	4.8	0.075	300	290	10.6	4000	0.6	N/A	2.	630
	2245	1500	17.	594	7260	46.3	100	7808	0.85	96.8	N/A	1.50	4.2	0.8	325	320	11.2	4000	1.0	27.8	1.9	720
	2345	1540	N/A	594	~7350	43.7	N/A	~7915	~1.31	95.0	1.47	N/A	4.	N/A	500	310	11.2	4200	0.8	46.8	2.	~780
11	0030	1510	15.4	590	7496	N/A	150	N/A	1.25	95.2	1.53	N/A	3.8	0.14	650	300	11.5	4500	1.0	54.3	2.	647
	0100	1515	N/A	N/A	N/A	N/A	250	N/A	N/A	N/A	N/A	N/A	N/A	N/A	400	330	11.5	4400	1.3	N/A	1.8	N/A
	0200	1550	15.4	590	7[illegible]	N/A	160	N/A	1.29	95.1	N/A	N/A	3.8	0.15	610	300	12.	5000	1.2	N/A	1.4	618
	0315	1515	N/A	N/A	N/.	33.	270	7706	2.3	91.4	1.36	1.69	N/A	N/A	550	310	12.	4500	2.5	40.8	1.3	N/A
	0400	1520	16.6	N/A	7098	N/A	160	N/A	1.26	95.2	N/A	N/A	4.5	0.24	1100	280	12.3	5000	3.0	N/A	1.3	638
	0510	1510	N/A	N/A	N/A	N/A	230	N/A	N/A	N/A	N/A	N/A	3.	N/A	900	280	12.	4600	2.	N/A	1.04	N/A
	0600	1530	16.5	N/A	7220	53.8	230	7610	1.33	92.8	N/A	N/A	3.	N/A	1700	230	12.	4500	1.3	38.	2.0	720
	0700	1501	17.7	595	7013	N/A	180	N/A	1.11	95.8	N/A	1.87	2.0	0.18	1500	240	12.1	5000	0.6	N/A	2.1	722
	0830	1540	N/A	595	N/A	52.1	220	7410	1.8	93.3	1.18	1.69	3.5	N/A	5000	270	12.4	4600	0.6	51.9	1.8	N/A
	0902	1540	N/A	650	6953	N/A	200	N/A	1.34	94.7	N/A	N/A	3.0	0.14	1700	270	12.5	4600	0.3	17.9	1.75	733
	1003	1540	16.	656	6818	50.6	N/A	7260	2.36	92.0	N/A	1.79	2.4	N/A	N/A	N/A	12.5	4600	0.3	47.7	1.8	823
	1051	1550	N/A	635	5818	N/A	255	N/A	2.26	92.3	N/A	1.83 (@ 1120)	2.5	0.34	2000	220	12.6	4600	0	N/A	1.55	N/A
	1200	1550	N/A	N/A	6883	55.	200	7297	1.61	94.3	1.04	N/A	5.	N/A	N/A	N/A	12.	4250	N/A	26.8	0.56	957
NR	1300	1520	N/A	N/A	N/A	N/A	200	N/A	N/A	N/A	N/A	N/A	1.5	N/A	1100	N/A	12.	4300	2.5	N/A	0.14	N/A
NR	1414	1510	N/A	467	N/A	41.7	150	6267	1.15	94.5	N/A	1.58	2.8	0.30	1500	270	11.8	3900	3.	11.6	0.07	N/A
NR	1520	1470	18.	525	5983	56.6	360	6376	2.53	89.3	3.32	1.58	3.	N/A	1500	N/A	11.2	3900	3.	7.8	0.03	990
NR	1630	1510	18.	550	7070	25.7	1000	N/A	7.8	68.5	N/A	1.52	6.8	0.04	900	270	10.8	4100	5.5	6.94	0.02	900
NR	1724	1493	19.	534	6890	54.4	1600	7264	14.2	40.8	3.	1.57	4.5	0.04	1000	270	11.	4100	5.	5.6	0.03	1110
NR	1817	Off	19.	Plug	N/A	N/A	N/A										N/A	3500	N/A	N/A	N/A	N/A

TABLE 6. (Continued) Page 4.

Day	Time	FBM S Input in Coal lb/hr	CBC Temp. °F	CBC NO ppm	CBC (CO₂ & SO₂) %	System C Burnup % Overall	CBC CO %	CBC HC ppm	FBM Limestone Add. Rate lb/hr	CBC Flue Gas lb/hr		CBC Flue Gas Sulfur lb/hr	CBC Coal Rate lb/hr	CBC Bed % S	CBC Fly Ash % S	CBC Fly Ash % Ca.	Concentration SO_2 Factor CBC/FBM
10	1243	25.4	2080	95	20.5	N/A	0.97	40	N/A	N/A		10.5	0	N/A	N/A	N/A	36
	1342	25.5	2120	550	23.	98.5	0.035	23	N/A	850		12.3	0	N/A	2.53	N/A	52
	1441	25.1	2080	1000	21.7	97.3	0.0035	0	0	1010		12.3	0	N/A	0.96	N/A	41
	1542	N/A	2060	1000	22.5	N/A	0.007	N/A	0	N/A		6.9	N/A	N/A	N/A	N/A	38
	1646	22.[illegible]	2020	920	>25.	96.3	0.08	5	30	734		11.8	0	3.35	0.77	N/A	47
	1758	23.	2020	660	>25.	97.7	0.18	28	29	1207		31.9	33.	N/A	1.08	22.3	53
	1857	24.3	2050	820	24.5	97.9	0.03	62	100	1120		29.	33.	2.01	1.22	N/A	34
	1953	N/A	2060	840	25.0	N/A	0.0105	42	0	N/A		24.7	N/A	N/A	N/A	N/A	89
	2045	N/A	2000	N/A	N/A	N/A	N/A	N/A	26	N/A		20.4	yes	N/A	N/A	N/A	N/A
	2130	N/A	1960	610	24.	N/A	0.35	12	26	N/A		14.8	0	N/A	N/A	N/A	200
	2245	26.7	[illegible]	460	24.	~97.8	0.26	3	26	800		16.1	11.8	N/A	1.19	N/A	190
	2345	26.7	1970	N/A	N/A	~97.4	N/A	3	39	~828	10	18.3	11.8	0.57	1.42	N/A	~136
11	0030	N/A	1940	N/A	23.	N/A	0.35	0.5	39	N/A	11	15.2	11.8	N/A	N/A	N/A	133
	0100	N/A	1970	250	N/A	N/A	N/A	4	43	N/A		13.7	11.8	N/A	N/A	N/A	72
	0200	N/A	1930	45	22.	N/A	0.02	12	43	N/A		10.2	16.7	N/A	N/A	N/A	88
	0315	~26.6	1900	N/A	N/A	~96.4	N/A	150	94	681		9.8	38.	N/A	1.2	N/A	49
	0400	N/A	1880	70	18.	N/A	0.5	250	94	N/A		9.8	20.	N/A	N/A	N/A	92
	0510	N/A	2000	60	N/A	N/A	N/A	100	94	N/A		7.8	40.	N/A	N/A	N/A	46
	0600	N/A	1890	190	N/A	~94.3	N/A	40	94	N/A		16.8	0	N/A	N/A	N/A	87
	0700	26.7	1940	310	25.	N/A	0.26	1.5	19	774		17.1	0	N/A	1.85	N/A	150
	0830	26.8	1970	N/A	N/A	96.3	N/A	N/A	19	777		15.4	0	6.53	1.62	10.73	82
	0902	29.2	1960	270	>25.	N/A	0.37	300	19	N/A		15.0	0	N/A	N/A	N/A	88
	1003	29.6	1880	0	N/A	~97.9	N/A	8	45	887		17.6	N/A	0.32	1.89	9.39	~61
	1051	28.6	1970	90	24.5	N/A	1.07	8	15	N/A		15.2	N/A	N/A	2.16	N/A	53
	1200	~28.6	1975	N/A	N/A	94.7	N/A	0	8	1017		6.3	0	0.36	N/A	N/A	28
NR	1300	N/A	2110	N/A	N/A	N/A	N/A	20	55	N/A		1.6	N/A	N/A	N/A	N/A	7
NR	1414	21.0	1960	880	16.5	98.0	0.006	47	0	~930		0.7	0	N/A	0.70	19.95	5
NR	1520	23.6	2030	N/A	N/A	98.7	N/A	150	0	1060		0.33	N/A	N/A	0.64	12.22	N/A
NR	1630	24.8	2060	1006	13.6	N/A	0.04	100	5	N/A		N/A	N/A	N/A	0.75	17.41	N/A
NR	1724	24.	2070	1080	15.	99.1	0.04	170	45	1176		N/A	N/A	N/A	0.84	13.5	N/A
NR	1817	N/A	1950	N/A	N/A	N/A	N/A	10	N/A	N/A		N/A	N/A	N/A	N/A	N/A	N/A

yielding an average of 92.7%. Continuous capture rate recording was not possible, since the single SO_2 analyzer must intermittently be switched to CBC gas analysis. The 98% carbon burnup goal (based on ash carbon analysis) was met or exceeded in 6 of the 34 burnup determinations.* 97% burnup was exceeded in 15 of the 34. In order to exceed 98%, it is necessary to avoid coal feed to the CBC, since the CBC combustion efficiency is rarely over 96%. In 10 of the 34, FBM fly ash carbon content** data appear to be in error based on ingredients rates and flue gas analysis. The carbon burnup vs. steam rate data are given below in Figure 19. The SO_2 concentration factor = 100 goal was met furing less than 50% of the test duration. During 98% carbon burnup conditions in the current apparatus, it is difficult to achieve as high as 3% SO_2 in the CBC regenerator off-gas (See Figure 20). The CBC air rate requirements for effective carbon burnup are the major factor in the problem. When coal was not being fed to the CBC, sulfur balances tended to be poor. This indicates that our measurement of a 3% SO_2 gas, using an 0.5% SO_2instrument with N_2 dilution, was unsatisfactory (the unit was later modified to 5.0% range).

The best compromise operating conditions in the apparatus used in this test appear to be exemplified by 2.4% SO_2, 97.9% burnup; and 1.6% SO_2, 98.2% burnup. These data are summarized in Table 7.

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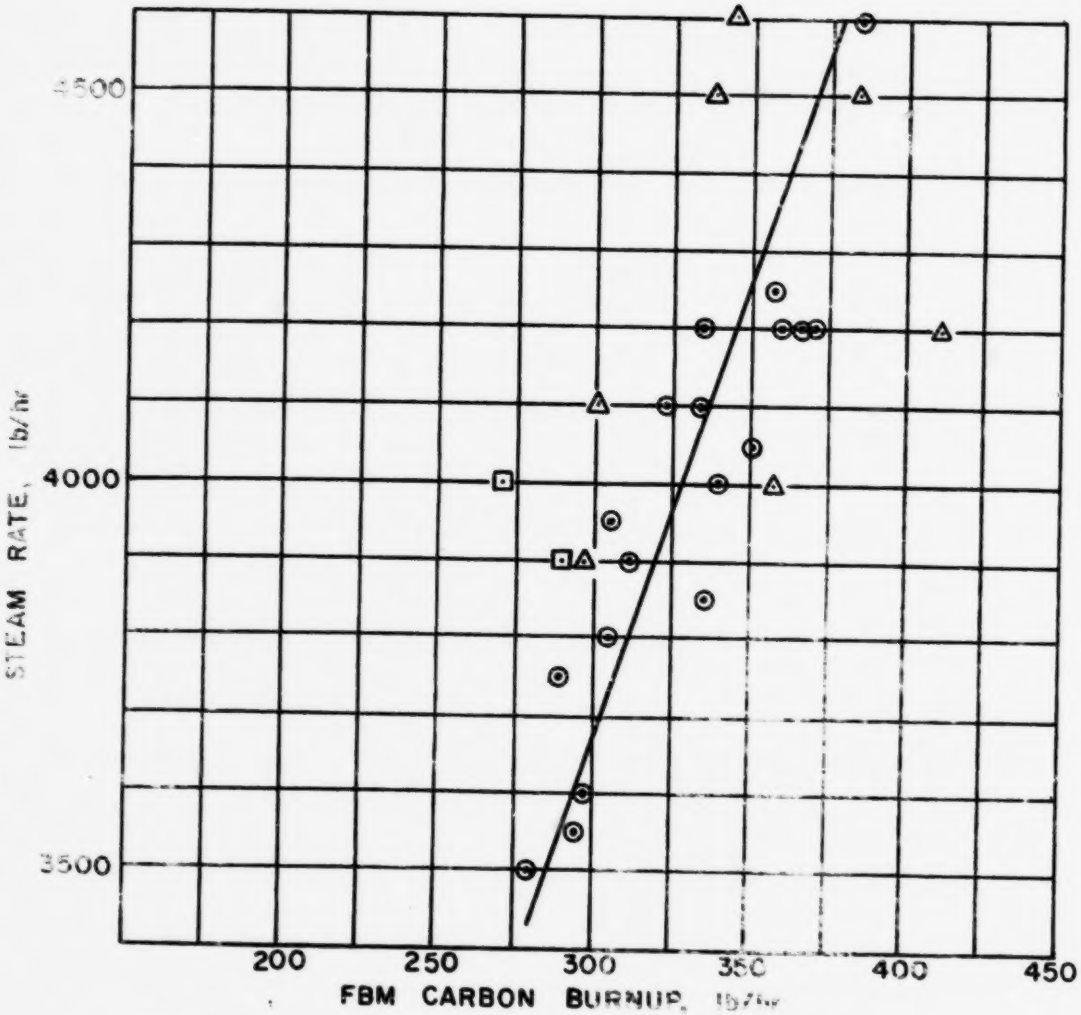
* Note that carbon burnups reported are probably conservative since no ash weighing system was provided. Final fly ash analytical results are listed in Appendix K.

** Fluctuations in FBM ash carbon content may be due to the current procedure of screwing the contents of the first (coarse) dust collector into the second (fine fraction) dust collector without a riffle box or other systematic mixing device.



⊡ COAL RATE MEASUREMENT IS QUESTIONED

△ BAD ASH % C

FIGURE 19 FBM STEAM RATE VS. CARBON BURNUP, RUN 168H

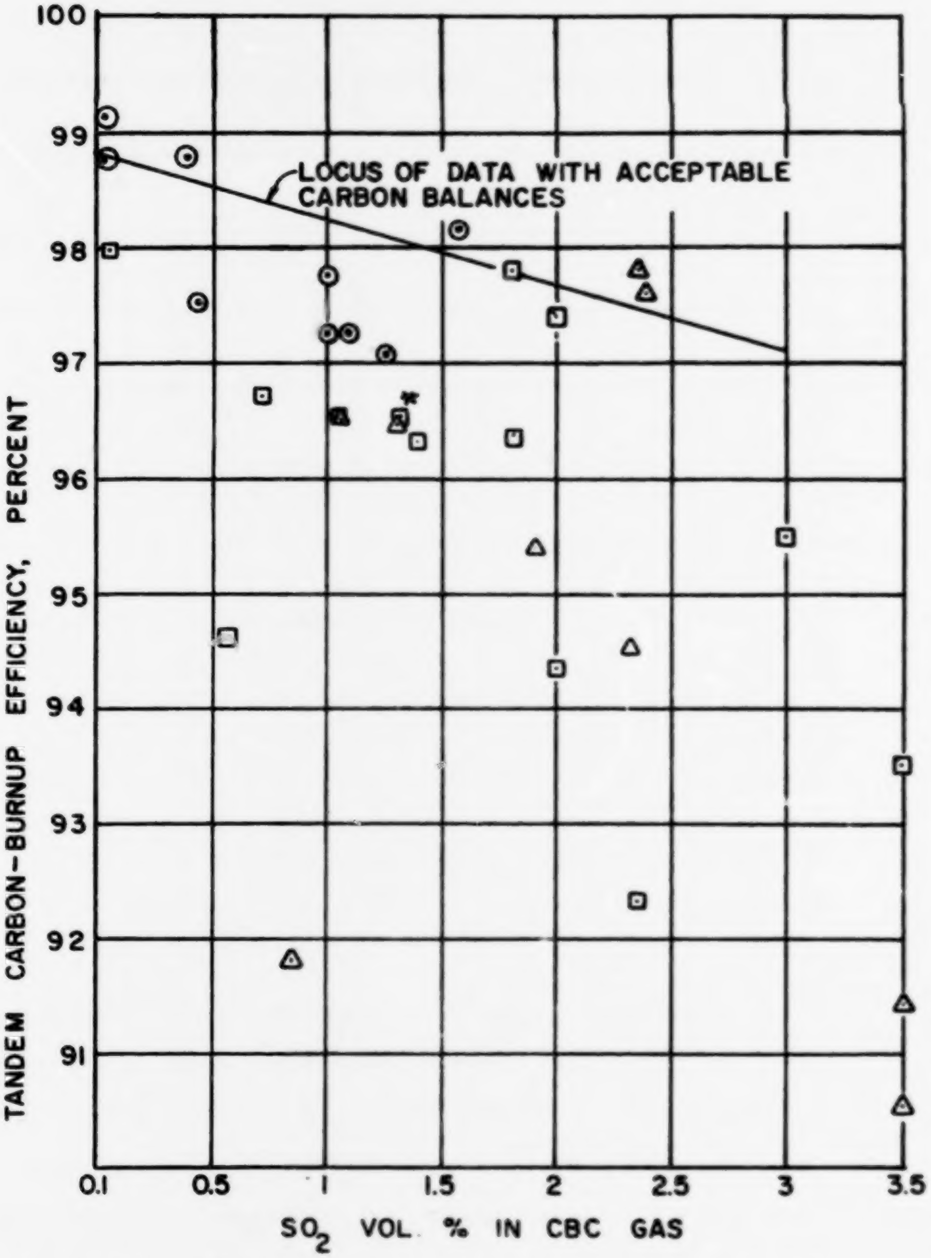


FIGURE 20. FBM-CBC SYSTEM CARBON BURNUP VS. CBC SO$_2$ OUTPUT, RUN 168H

⊡ BAD ASH C ANALYSES

△ COAL FEED TO CBC

TABLE 7. SUMMARY OF BURNUP DATA, RUN 168H

Point No.**	CBC O_2 %	CBC SO_2 %	C_B Overall %	CBC T, °F	S Capture in FBM %	Remarks
30B	5.	0.03	99.1	2070	N/A	Non-regenerating
28B	3.	0.03	98.7	2030	89.3	Non-regenerating
27B	3.	0.07	98.0	1960	94.5	Non-regenerating
19A	2.4	0.4	97.5	2020	90.1	
21A	4.5	0.4	98.7	1930	93.8	
25B	N/A	0.56	94.7	1970	94.3	C imbalance***
7A	1.1	0.72	96.7	1840	93.3	C imbalance
35A	2	0.84	91.8	2010	94.5	Coal to CBC
3A	3.5	1.0	97.7	2000	91.6	
39A	3.5	1.05	96.6	2010	92.7	Coal to CBC
37A	3.	1.1	96.6	1990	91.6	Coal to CBC
3B	1.5	1.1	97.3	2080	91.5	
20A	2.6	1.25	97.1	1990	84.6	
16B	2.5	1.3	96.4	1900	91.4	Coal to CBC
2B	1.8	1.3	98.5*	2120	92.1	C imbalance
5B	0.5	1.45	96.3	2020	89.3	C imbalance
28A	1.3	1.6	98.2	2020	92.1	
23B	0.3	1.8	97.9	1980	92.	C imbalance
21B	0.6	1.8	96.3	1970	93.3	C imbalance
11B	1.0	1.9	97.8*	1930	96.8	Coal to CBC
12B	0.8	2.0	97.4	1970	95.	Coal to CBC
19B	1.3	2.0	94.3	1890	92.8	C, O imbalances
7B	0.4	2.35	97.9	2050	91.6	Coal to CBC
5A	2.0	2.4	92.3	1955	93.1	C imbalance
14A	2.0	2.4	94.5	1840	N/A	Coal to CBC
6B	0.1	2.4	97.7	2020	90.8	Coal to CBC
29A	1.2	3.0	95.6	1980	91.4	C imbalance
23A	0.	4.	93.6	1930	88.6	C imbalance
24A	0.	4.	91.4	1920	89.9	Coal to CBC
38A	3.	1.2	~89.	2000	92.2	C, O imbalances
13A	2.5	N/A	~93.3	1910	90.8	Coal to CBC
17A	3.	1.0	97.3	2000	94.	
25A	1.2	4.	90.5	1980	91.[illegible]	Coal to CBC
36A	3.2	1.2	88.	1980	92.	Coal to CBC

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[*] These carbon burnup values based on flue gas analysis were not corroborated by heat balance calculation (See Section 6.3.2.2).

[**] See Table 6

[***] Flue gas rate and analysis not in agreement with ash-based burnup calculation.

Based on the test results, process equipment redesign was undertaken, based on a 3-reactor SO_2 Acceptor Process concept: (1) FBM as currently set up; (2) an auxiliary fluidized regenerator of 0.6 ft^2 cross section, burning coal and regenerating bed material pumped from the FBM; (3) a conventional CBC, essentially as currently set up, burning fly ash from (1) or from both (1) and (2). The bed temperatures of (2) and (3) are essentially equal. Regenerated bed material from (2) returns by fluidized flow or gravity to either (1) or (3).

One possible way to promote carbon burnup under regenerating conditions is to operate the CBC at higher temperatures*. For example, at 2120°F, 98.5% burnup was achieved at 1.3% SO_2 and 1.8% O_2 (see point 2B on Table 7). At 2020°F, 98.2% burnup was achieved at 1.3% O_2 and 1.6% SO_2 (see point 28A). CBC temperatures above 2120°F were avoided in this test (with a refractory lined CBC) from fear of slagging. However, the ash fusion behavior in a large excess of lime environment, when no coal is fed has not been studied, and this fear may be unfounded. The 3-reactor concept, of course, obviates excessively high CBC temperatures. The boiler system was operated for over three days during run 168H, with one interruption of about 2 hours' duration. There were two minor upsets of 2 and 15 minutes when steam production rate was reduced but did not cease. In each instance of SO_2 capture below 90%, temporary limestone feed disorders were the sole cause. Bed material for analysis is removed at about 10 lb/hr; this, plus attrition, must be made up intermittently as needed to trim

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* Another way is to reduce gas velocity, by building a CBC of larger grid area.

FBM bed level, bed temperature, and capture percentage. Based on the test results (see limestone rate data in Table 6) a limestone requirement for this process of about 5% of coal rate appears reasonable. The Powhatan coal did not appear to add oversize inert matter (rocks) to the bed. Calculations based on fly ash calcium analyses tended to confirm the 5% value. The coal particle size analysis for this test is shown in Figure 21.

6.3.2.2 Gaseous Emissions (Run 168H)

The available FBM SO_2 emission vs. bed composition data are collected in Figure 22. There is considerable data scatter due to the combined effects of temperature, O_2 level, and previous history of the lime bed. While there is an incentive to keep bed sulfur below 3.5% to guarantee good capture, it is interesting to note that good capture can still be obtained at sulfur levels above 7% with continuous passage of bed material through a regenerator. This capability is not observed in batch experiments where bed sulfur continually rises, and is probably due to surface or "superficial" regeneration of particles. The small particles having higher surface area/volume ratio, are more completely regenerated and are more fully activated for subsequent SO_2 capture. The larger particles retain a core of higher $CaSO_4$ content* after passing through the regenerator, and bias the overall analytical result, so that a 7% sulfur bed, after undergoing heavy regeneration, can [illegible] as a 3% sulfur bed would after undergoing a lower level of regeneration. Evidence was presented in Section 6.2, page 6-20 for lower S/Ca ratio on the surface of particles

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* See discussion of the [illegible] concept in Section 6.2, page 6-20.

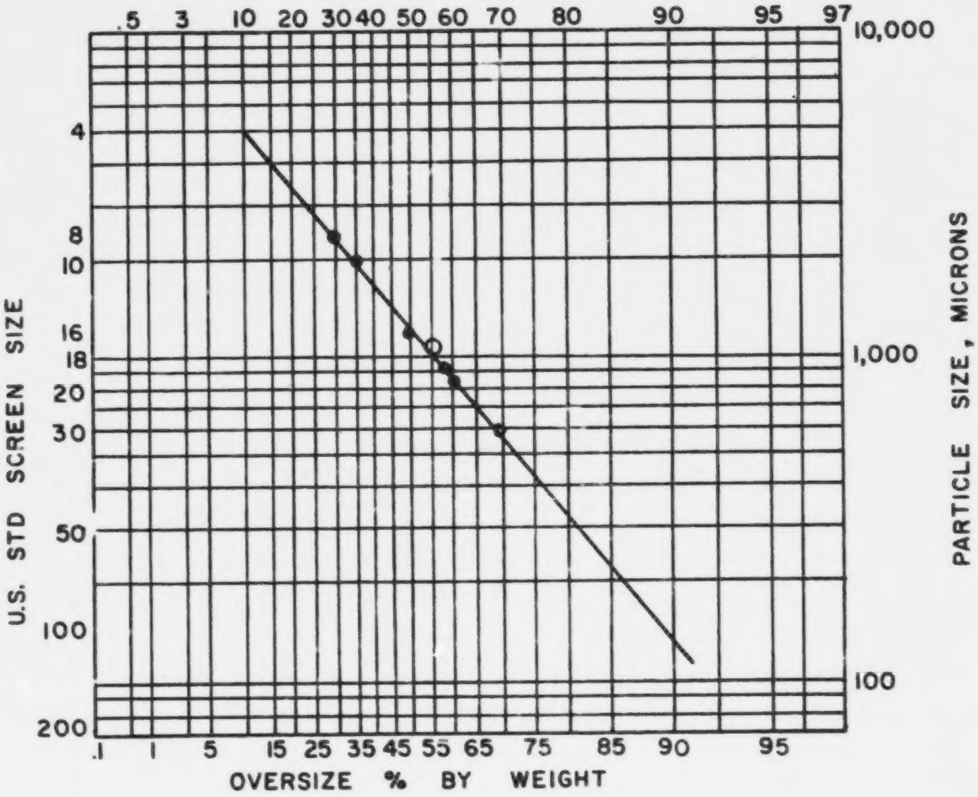

FIGURE 21. COAL PARTICLE SIZE ANALYSIS, RUN 168H

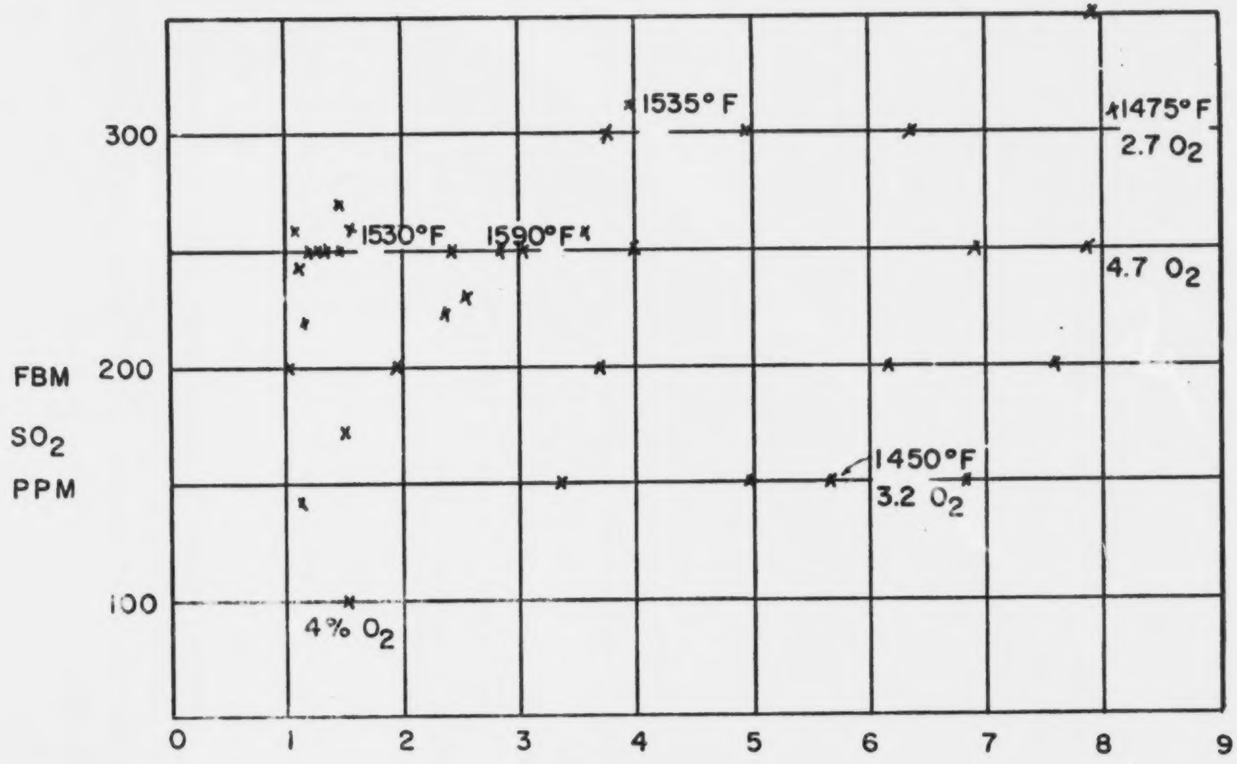

FIGURE 22. FBM FLUE GAS SO_2 CONCENTRATION VS BED SULFUR CONTENT – RUN 168H

in an aged bed. A few selected bed samples were screened and the screen fractions analyzed for S, but results were inconclusive.

At one point in the test, salt feed to the FBM was performed (see Table 6). The immediate effect was an increase in coal combustion efficiency in the FBM causing the CBC temperature to drop since it was suddenly starved for fuel (and its O_2 output to increase) and the SO_2 output of the FBM was immediately cut by 47%, confirming the earlier FBC test results with salt feed.

The available FBM NO emission vs. bed composition data are collected in Figure 23. By operating at 6% sulfur, rather than 1% sulfur, a 10% reduction in FBM NO emission is predicted. Note from Table 6, however, that a typical split of NO emissions is 2/3 FBM, 1/3 CBC due to the higher temperature and O_2 level of the CBC. Several possible explanations for the slight but noticeable effect of bed sulfur level and NO production exist:

a. The effect is real, and possibly due to oxidation of nitrogenous matter in coal by $CaSO_4$ in a mild fashion yielding N_2 rather than NO, in relatively oxygen-deficient regions of the bed near the coal feed points.

b. The effect is real and possibly due to stickiness of $CaSO_4$ laden particles to coke balls thereby covering the balls and affecting coal activity. Low coal activity would reduce "hot spot" NO generation. Such coke balls are frequently observed with lime particle coats when large lump size coal (3/4") is fed. The formation of a low melting CaS - $CaSO_3$ - $CaSO_4$ transient eutectic appears possible.

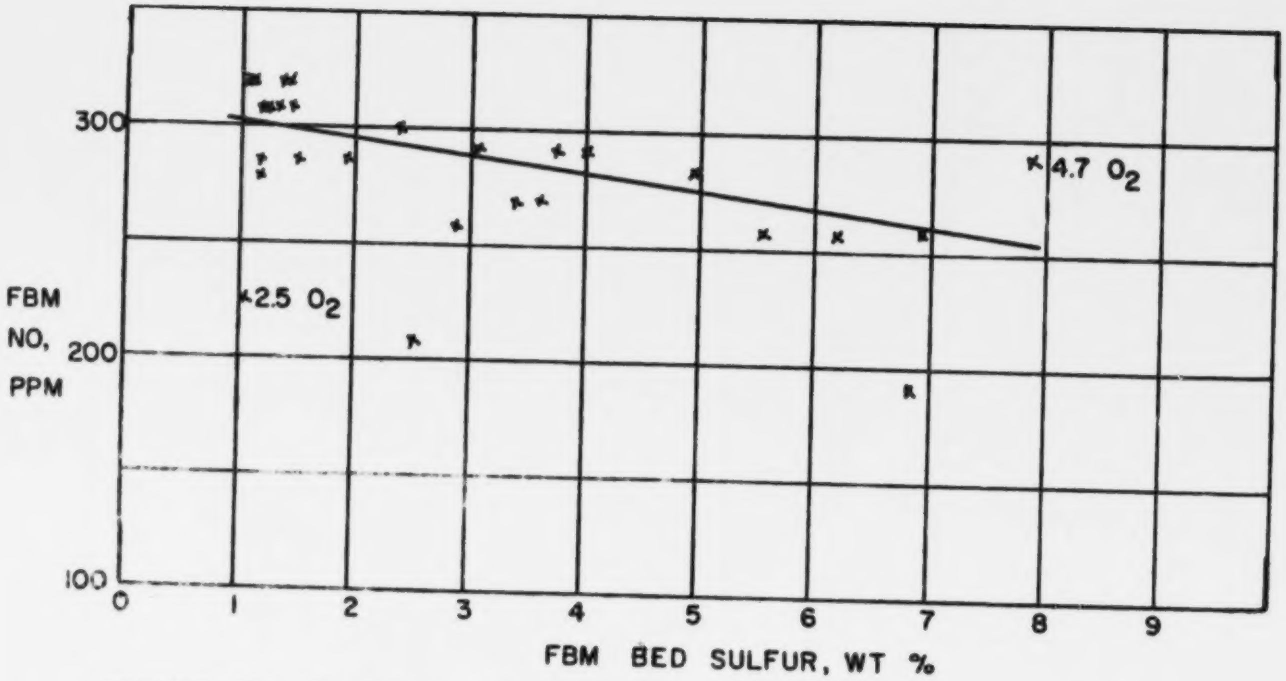

FIGURE 23. FBM FLUE GAS NO CONCENTRATION VS BED SULFUR CONTENT RUN - 168H

c. The apparent effect is an artifact due to lower bed mean particle size at higher $CaSO_4$ content and particle density. A bed of smaller particles could lead to greater temperature uniformity and reduction of hot spots believed to generate excess NO. A set of experiments in the FBC using coarse and fine limestone fractions could be expected to elucidate this possibility. Experiments on NO reduction had been scheduled as part of this contract but were cancelled when the long duration run became the priority item.

Condensed data on CO and HC emissions in this run are in Appendix I.

6.3.2.3 Boiler System Heat Balance

Heat balance data for several selected conditions during this run are collected in Table 8. Only those conditions in which carbon-burnup efficiency as calculated from ash carbon analyses and/or flue gas analyses is above 96.5% are treated here. When coal is added to the CBC, heat balances tend to be poorer. The reaction heats of calcining fresh limestone, and capture and desorption of SO_2 are accounted for under a term, "net $CaSO_4$ formation." The unaccounted heat is correlated with apparent thermal efficiency of the system in Figure 24. This indicates that the major measurement errors are occurring in the ingredients rates and steam output, rather than in the individual loss terms. Recorders for coal rates and air rates would have provided a more accurate record than data logged by operators.

6.3.2.4 Bed Specific Gravity

Since bed sulfur content seemed to have a noticeable effect on pollutants release rate, it was hoped that bed specific gravity could be used to provide a quick, on-line

TABLE 8. HEAT BALANCE DATA RUN 108B

Day Time	11 1724	11 1530	11 1414	9 1152	9 1404	9 0007	8 2015	9 0940	10 0930	10 1441	10 0730	9 1312	10 1343	9 2130	11 1003	10 2343	10 2345	10 1857	10 1758	AVG
FBM bed temperature, °F	1490	1470	1510	1475	1485	1550	1540	1480	1525	1530	1530	1480	1530	1540	1540	1500	1540	1480	1400	
SO_2 Capture, %	90	89.3	94.5	80.1	93.8	93.3	91.6	94	92.7	91.5	91.6	88.0	93.1	93.1	92	96.8	95	91.6	90.8	
Carbon Burnup, %	99.1	98.7	94	97.5	98.7	96.7	97.7	97.3	98.6	97.3	96.6	97.1	99.4	95.2	97.8	95.4	97.4	97.9	97.8	
CBC SO_2, %	0.03	0.03	0.07	9.4	0.4	0.72	1.0	1.0	1.68	1.1	1.1	1.25	1.3	1.6	1.8	1.9	2	2.35	2.4	
Total preheated air, lb/hr	8000	8400	8600	* ~7800	7000	6400	6400	7000	7700	7500	7000	6900	7500	7500	7400	7600	7700**	7700	7600	
Total preheated air, KBtu/hr	475	393	392	451	503	444	459	450	477	430	500	501	483	376	488	434	457	438	451	
CBC Air, lb/hr	1110	990	990	1040	730	730	730	1130	890	980	890	880	950	770	910*	720	780	1080	1090	
Coal Heating value, KBtu/hr	6085	6087	6100*	7250	7129**	6722**	6095	7070	8107	7289	8048	6848	7420	7108	8881*	7936	7936	7485	7103	
Net $CaSO_4$ Formation, KBtu/hr	28	N/A	125	103	118	97	50	60	N/A	N/A	N/A	57	N/A	N/A	N/A	N/A	17	-103	-98	
Total input, KBtu/hr	7402	7255	6625	7811	7741	7221	6580	7590	9043	7720	8553	7300	7900	7763	9033	8374	8394	7826	7455	
Steam absorption, KBtu/hr	4161	3950	3959	4153	3880	3806	3568	3951	4255	4183	4233	4052	4103	4000	4000	4052	4255	4052	4068*	
Circ. H_2O absorption KBtu/hr	1633	1538	1530	1673	1530	1661	1417	1464	1551	1552	1636	1518	1553	1560	1563	1621	1585	1612	1634	
Total (H_2O+St.)	5794	5497	5487	5826	5518	5467	4983	5415	5806	5655	5611	5570	5714	5600	6352	5673	5840	5664	5626	
% of coal value	92*	90.1	90*	80.3	77.4	81	81.7	76.5	71.1	77.5	73.4	80.3	76.9	78.6	73.0	71.6	73.7	75.6	79.3	
Flue Gas loss, KBtu/hr	941	605	625	903	815	740	740	834	932	882	858	782	883	1083	848	885	900**	814	937	
Blowdown allowance KBtu/hr	21	20	20	24	30	22	21	23	25	24	25	24	21	21	24	24	28	24	23	88
Hydrocarbon loss, KBtu/hr	117	115	110	31	18	64	31	19	64	14	30	22	21	3	172	30	48	34	48	186
CO loss, KBtu/hr	9	11	79	372	232	188	180	251	84	87	37	362	48	87	103	236	58	33	53	188
CBC Carbon loss KBtu/hr	51	73	93	145	74	174	132	288	235	157	218	124	385	102	143	383	180	124	122	
% of coal value	0.73	1.08	1.6	2.0	1.04	2.59	1.8	3.7	2.8	2.1	2.7	1.8	3.56	1.4	1.60	4.5	2	1.64	1.7	
Hot Fly Ash loss KBtu/hr	23	19	15	24	22	24	18	23	33	30	22	14	34	36	33	23	20	20	20	
FBM collector dust carbon loss	172	146	52	99	116	103	56	103	56	34	73	20	87	87	108	85	73	58	73	
Nitric oxide loss KBtu/hr	N/A	3	N/A	4	N/A	3	3	4	4	4	4	4	N/A	N/A	3	4	4	N/A	4	
Lift air loss KBtu/hr	N/A	36	N/A	3*	38	37	39	34	34	35	31	36	40	64	38	32	31	38	37	
Radiation & unaccounted	N/A	276	N/A	N/A	388	N/A	N/A	40	902	400	790	130	230	105	881	842	781	470	61	
Total Losses	1283	1368	1001	1628	1781	1354	1210	1716	2359	1640	2122	1420	1712	1686	2233	2267	3067	1734	1378	

←Coal to CBC→

Coal to CBC

←Coal to CBC → → →

* Rate Questioned

** Rate Estimated

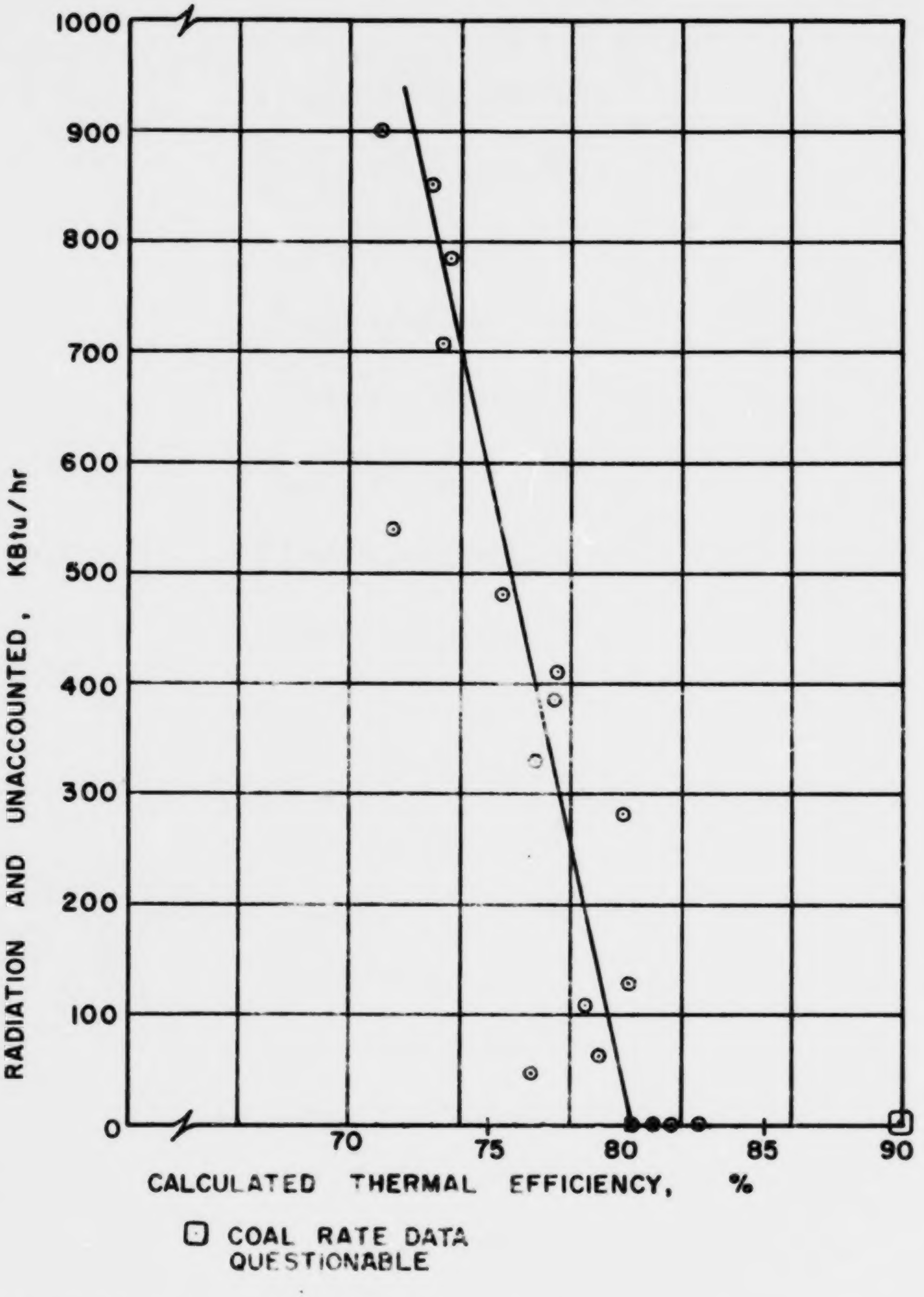

FIGURE 24. HEAT BALANCE - RUN 168 H

estimate of sulfur content without waiting for chemical analysis. Specific gravities (poured without tapping) were measured in a 100 cm^3 graduate*. Data are collected in Appendix J. Although the highest bed densities occur at the higher sulfur contents, data scatter is excessive and other factors (particle size, shape, coal ash ingredient contaminants) must also affect bed density. In most cases, CBC beds are about 2% less dense than FBM beds, in line with the effect of regeneration.

Bed particle size distributions for this run are described in Section 6.4.

6.3.2.5 Sulfur Analyses

Fly ash and bed material sulfur analyses were performed on many of the FBM and CBC samples (see Table 6). Most fly ash samples are below 2% S. The highest FBM fly ash S% is 1.89 (average 1.69). When coal is fed to the CBC, the CBC fly ash tends to be high in both C and S, implying that some S is being eliminated from the system, tied up in pyrolized coal material rather than as $CaSO_4$. Under regenerating conditions, without coal feed to the CBC, its ash contains less S than the FBM fly ash input; i.e., the calcium sulfate content of the ash is being "regenerated" along with the bed material. The average CBC ash is 1.53 %S, 10% less than the FBM average. For comparison, the fly ash S contents of the FBC coal-burning nonregeneration tests previously discussed were all over 2.2% tending to indicate that fly ash from the regenerative FBM operation contains surface material abraded from the low-sulfur surface layer of the partly regenerated particles.

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* Ash particle densities can be measured using water displacement, but lime particles react with water.

The sulfur balance data for this run are presented in Appendix F, Table 2. The imbalance is 15%. The principal error is believed to occur in the measurement of CBC SO_2 level, using the 0.5% instrument with N_2 dilution (see Section 5.5). A 17% error in measuring a 3% SO_2 stream by this technique appears reasonable; thus it is assumed that the measured 92.7% S capture in the FBM is accurate. This instrument has since been converted to triple range operation: 0-900, 0-5000 ppm and 0-5% SO_2. A separate continuous 0 to 10% SO_2 regenerator flue gas recording analyzer is required. This would have permitted on-line FBM and REG/CBC SO_2 recording without interruptions.

6.3.3 FBM/CBC Run with Coal Feed to CBC (Run 169H)

Following completion of the FBM/CBC long duration run, another test was made in which FBM fly ash was discarded and the CBC run on coal only. CBC temperatures were in the 1950-1990°F range. Six levels of CBC O_2 were studied between 0.4 and 3.0 percent. Used lime bed material from the previous run (168H) continued to give good SO_2 capture in the FBM.

The carbon combustion efficiency of the CBC burning coal only during test 169H is plotted versus the oxygen content of the CBC off gas in Figure 25. These points are shown as circles. It may be seen that there is a poor correlation between O_2 and carbon combustion-efficiency. In experiments described in Reference 1 a similar burnup efficiency was achieved at a far higher O_2 level (Test C-319-5, 95.5% efficient, 9.8% O_2, 1980°F).

Data from Test 168H is also potted on Figure 25. For this test the carbon burnup efficiency values are for the tandem system, i.e. CE = (Carbon burned/Carbon fed

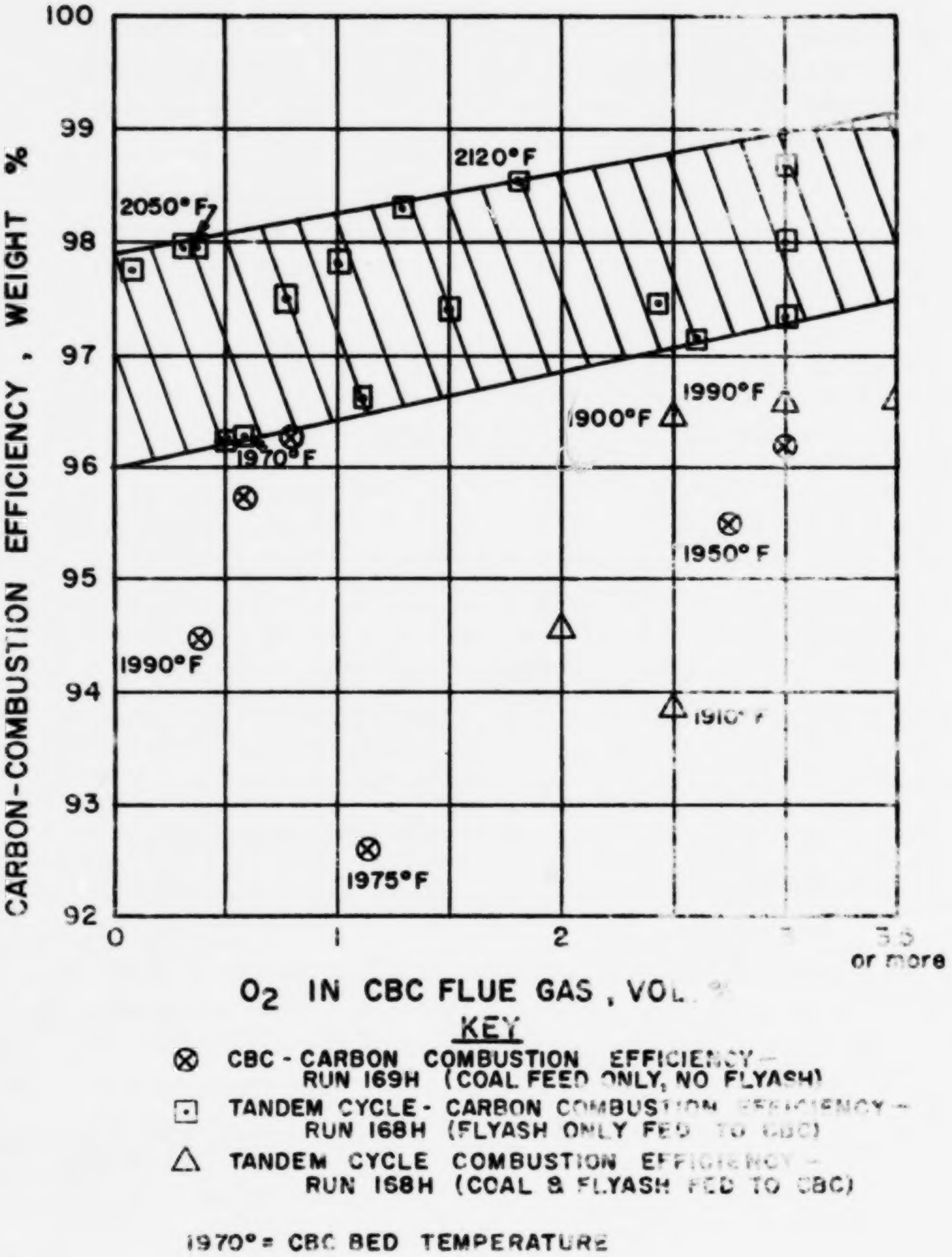

FIG. 25.- CBC CARBON BURNUP AND TANDEM CYCLE BURNUP VS. OXYGEN LEVEL

to FBM and CBC) x 100. When fly ash was the sole fuel to the CBC (points shown as squares in Figure 25) the combustion efficiency was relatively high. When coal was added to the CBC (points shown as triangles) the systems carbon combustion efficiency was lower.

The band shown on Figure 25 shows the data scatter which probably results from differences in factors such as fly ash feed particle size, carbon content and feed rate.

Sulfur balances with coal feed appear good (see Table 9). Unlike the situation with fly ash feed for fuel, good regeneration with coal feed can be achieved at any CBC flue gas O_2 level between 0% and 3%.

The goal of CBC flue gas being 1/11 of the total FBM plus CBC flow (or less) was not achieved, limiting CBC SO_2 level to less than 3% in several cases. As previously, the $CaCO_3$ makeup needs are modest (20 lb/hr including bed sample withdrawals). The CO analayzer was inoperative for this test. The desired design data for the 3-reactor concept (see below) were obtained from this test.

6.4 Three-Zone SO_2 Acceptor Process MK II Concept (FBM/CBC/REG)

6.4.1 Background

The difficulties encountered in attempts to continuously regenerate sulfated lime beds in the existing CBC have been described. The CBC flue gas SO_2 levels were usually too low. The existing CBC was designed to achieve high carbon burnup efficiency in a high-velocity apparatus using high excess air levels. Due to the laboratory structural layout, it was not feasible to rebuild the CBC to utilize low-velocity combustion at lower excess air levels to achieve the same burnup efficiency. Another alternative which could not be pursued was

TABLE 9. FBM-CBC RUN WITH COAL FEED TO CBC (Run 169H)

CONDITION NO.	1	2	3	4	5	6
FBM: O_2, %	8.5	8.5	8.5	7.8	4.2	3.8
HC, ppm	120	120	120	130	800	800
SO_2, ppm	25	25	25	150	80	80
NO, ppm	330	330	330	290	250	250
CO_2, %	10.5	10.5	10.5	9	[illegible]	11.2
Steam, lb/hr	3700	3700	3700	3700	[illegible]	3450
Fly Ash C, %	45.5	48.8	47.5	51.3	[illegible]	N/A
Air Rate, lb/hr	9080	9080	9080	8950	6114	6314
CBC: T, °F	1990	1990	1990	1970	1950	1950
O_2, %	0.4	0.6	0.8	1.2-1.8	2.75	3.0
*CO_2, meter (%)	>25	>25	>25	21.5	19.5	19.5
HC, ppm	60	60	60	6	8	10
SO_2, %	3.6	3.6	3.6	2.49	2.0	2.95
NO, ppm	390	390	390	540	580	580
Air Rate, lb/hr	955	955	955	928	897	907
S Emission, lb/hr	38	38	38	25.4	19.8	29.5
Fly Ash C, %	24.3	19.8	17.9	30.4	20.7	18.2
C Burnup, % CBC only	94.5	95.8	96.3	[illegible]	95.6	96.2
Coal Input Rates lb/hr: FBM	620	620	620	[illegible]	510	540
Coal Input Rates lb/hr: CBC	82	82	82	[illegible]	82	82
Bed Circulation Rate, lb/hr	380	380	380	[illegible]	[illegible]	880
Sulfur Input Rates lb/hr: FBM	28	28	28	28	23.4	23.4
Sulfur Input Rates lb/hr: CBC	3.7	3.7	3.7	[illegible]	3.7	3.7

Bed Material Bulk Density 73.5 lb/ft³
Limestone Makeup 154 lb/8 hrs.

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*Includes SO_2 response
(thermal conductimetric determination of CO_2 [illegible]

Note: N/A = not available

increasing the bed level of the CBC. It is unfortunate that this course could not have been pursued. The high excess air principle was found to be incompatible with bed regeneration when fly ash was the sole fuel. The use of CBC coal feed to achieve rapid regeneration was found to lower carbon burnup efficiency. PER reported to EPA that EPA's goals (high carbon burnup, high regenerator flue gas SO_2 concentration, high FBM flue gas desulfurization) might be met most expeditiously in a 3-zone apparatus in which the high-temperature functions (burnup and regeneration) were physically separated.

6.4.2 Extended Run Test Operation (Run 171H)

Following construction and shakedown testing of the three-reactor system, a long duration test of the integrated FBM/CBC/REG pilot plant scale boiler system in the SO_2 Acceptor Process MK II (three-cell) mode, was conducted over an 8-day period. One hundred fifty-six hours of boiler operation were logged. Coal fed was 37 tons. Average regenerator flue gas SO_2 concentration was 4.1% (volume) or 13.5 lb/hour, (25 determinations), as determined by IR and wet assay techniques. The FBM SO_2 capture and CBC burnup efficiency data are discussed below. Some down time of the system resulted from (a) the developmental nature of the regenerator (it was an appendage to the FBM) and (b) lack of maintenance on the FBM/CBC components due to budget constraints. See Test Log in Table 10.

A 6.1% ash, 3.3% S coal (Powhatan) was used during most of the test and its average carbon burnup exceeded 99%; the highest CBC output carbon level was 11.9%*

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* C level in CBC dust collector collected fly ash.

TABLE 10

FBM/[illegible]/[illegible]. SYSTEM LOG - Run 171B

FBM

CBC :

Day	Time	T, °F	Bed Level In	Coal Rate, lb/hr	Air Rate lb/hr	Fly Ash Carbon %	Flue Gas SO_2 ppm	Flue Gas Rate lb/hr	Bed S %	Bed Ca %	O_2 %	CO %	HC ppm	CO_2 %	Steam lb/hr	O_2 %	Ash C, %	SO_2 ppm	Air Rate lb/hr	C ** Burnup % Overall	T, °F	CO ppm	CO_2 %	HC ppm
7/12	1000	1500	N/A	N/A	N/A	N/A	820	N/A	N/A	N/A	3.5	0.22	1600	17.	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	1200	1560	N/A	445	N/A	N/A	600	N/A	N/A	N/A	7	0.06	1400	15.	3200	N/A	N/A	N/A	N/A	N/A	1920	1300	N/A	N/A
	1400	1550	N/A	550	N/A	N/A	320	N/A	N/A	N/A	4.	N/A	1500	N/A	3700	N/A	N/A	820	N/A	N/A	1950	1200	N/A	N/A
	1600	1630	N/A	510	N/A	N/A	410	N/A	N/A	N/A	6.5	0.16	1400	9.	3200	N/A	N/A	N/A	N/A	N/A	2010	1100	N/A	N/A
	1800	1525	N/A	425	4700	49.6	370	5090	3.6	N/A	3.5"	0.12	1000	8.5	3000	5.	N/A	N/A	1176	N/A	1970	400	N/A	N/A
	2000	1480	N/A	405	[illegible]	39.7	310	4980	3.61	13.3	4.6	N/A	1000	8.3	3100	3.	0.	600	1167	99.	1960	360	N/A	N/A
7/12	2200	1480	N/A	521	4860	N/A	200	4920	4.51	N/A	4.3	0.22	1000	8.2	3200	N/A	1.	2500	1264	99.	1940	280	N/A	N/A
	2400	1500	N/A	476	[illegible]	36.5	N/A	5030	N/A	N/A	4.2	N/A	N/A	N/A	3100	2.	N/A	2400	1035	N/A	1920	900	15.	100
7/13	0200	1540	N/A	[illegible]	4650	53.5	N/A	5020	3.48	N/A	2.6	N/A	N/A	N/A	3200	5.	1.	1000	1114	99.	2010	80	N/A	N/A
	0400	1480	N/A	465	N/A	N/A	220	N/A	4.86	N/A	3.0	N/A	N/A	9.5	3100	6.	1.	240	N/A	99.	1920	60	N/A	N/A
	0600	1520	16	481	4770	N/A	220	5150	N/A	N/A	3.0	0.2	1900	9.3	3100	N/A	N/A	270	990	N/A	1980	50	14.	0
	0800-0845	1540	16	451	5258	49.9 (845)	550	5594	3.56	N/A	3.0	0.35	1500	9.0	3100	N/A	1.	410	975	N/A	1950	30	N/A	N/A
	1000	1480	N/A	409	N/A	36.2	460	N/A	3.57	23.22	2.6	0.45	1700	9	3100	N/A	1.8	370	N/A	99.	1960	100	12.5	90
	1200	1460	N/A	445	N/A	27.5	280	N/A	3.92	N/A	3.5	0.4	1500	9.1	3100	7.5	0.7	1200	N/A	99.	1960	50	N/A	N/A
	1400	1420	N/A	475	N/A	N/A	200	N/A	N/A	N/A	4	0.23	1000	9	3200	N/A	N/A	5000	N/A	N/A	1910	50	N/A	N/A
	1600	1500	17	460	5650	N/A	N/A	6010	N/A	N/A	3.3	N/A	N/A	N/A	3250	5.	N/A	500	1245	N/A	1970	100	13.	50
	1800	1460	N/A	497	N/A	N/A	200	N/A	N/A	N/A	3.3	0.45	1600	9.3	3250	N/A	N/A	100	N/A	N/A	1970	55	N/A	N/A
	2000	1460	N/A	474	N/A	21.8	N/A	N/A	4.74	N/A	3.4	N/A	N/A	N/A	3300	7.	0.5	1000	N/A	99.	1960	60	13.	48
7/13	2200	1480	17	515	6220	27.3	360	6600	3.74	N/A	3.4	0.42	1550	10.1	3300	7.	1.	190	1315	99.	1980	50	N/A	N/A
	2400	1540	N/A	499	N/A	N/A	350	N/A	N/A	N/A	2.9	0.5	1800	11	3300	N/A	N/A	N/A	1315	N/A	1980	N/A	N/A	N/A
7/14	0050	N/A	N/A	N/A	N/A	N/A	240	N/A	N/A	N/A	5	0.15	N/A	10.7	3500	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	0400	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	1800	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	4.2	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	2010	1440	N/A	500	N/A	N/A	160	~ 6600	N/A	N/A	6.5	0.12	270	12.	3050	N/A	N/A	700	N/A	N/A	N/A	90	N/A	N/A
7/14	2200	1510	N/A	505	N/A	N/A	190	~ 6600	1.32	22.5[illegible]	6	0.11	300	12	2950	5	N/A	110	N/A	N/A	1960	150	12.	150
	2400	1530	N/A	518	N/A	[illegible]	225	~ 6600	N/A	N/A	5.5	0.13	450	12.2	3350	N/A	3.9	130	N/A	99.	2010	60	N/A	N/A
7/15	0200	1450	N/A	[illegible]	N/A	N/A	270	~ 6600	3.48	N/A	6.3	0.1	250	12.	3350	7	N/A	280	N/A	N/A	1920	50	13.	85
	0400-0545	1440	N/A	600	N/A	N/A	N/A	~ 6600	2.98	N/A	6.5	N/A	N/A	N/A	3350	N/A	N/A	N/A	N/A	N/A	1930	N/A	N/A	N/A
	0600-0745	1470	N/A	508	N/A	N/A	240	~ 6600	3.4	N/A	8	N/A	N/A	8	3350	N/A	N/A	N/A	N/A	N/A	1800	N/A	N/A	N/A
	0800-0845	1500	N/A	500	[illegible]	50.4	270	7120	3.7	N/A	4.5	0.06	750	9.	3150	N/A	11.95	N/A	1300	99.0	1870	N/A	N/A	N/A
	1000	1460	[illegible]	560	N/A	50.6	320	~ 7120	4.96	17.37	3.6	0.2	600	9.5	3400	4	2.6	320	N/A	99	2020	100	11.5	[illegible]
	1200	1500	N/A	600	N/A	46.8	600	~ 7120	3.22	N/A	5.7	0.4	2500*	6	3400	N/A	3.9	N/A	N/A	99.	2030	N/A	N/A	N/A
	1400	1510	N/A	N/A	N/A	N/A	500	~ 7120	N/A	N/A	8	0.3	1600	6	3400	N/A	N/A	200	N/A	N/A	2000	N/A	N/A	N/A
	1625	N/A	N/A	534	N/A	N/A	[illegible]	~ 7120	N/A	N/A	N/A	N/A	*	8	35[illegible]	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
15	1800	N/A	N/A	N/A	N/A	N/A	[illegible]	N/A	N/A	N/A	N/A	N/A	N/A	N/A		N/A	N/A	N/A	N/A	N/A	N/A		N/A	N/A

* Coal Feed Plug

** System Burnup

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TABLE 10 (CONTINUED)

Day	Time	System S Input In Coal, lb/hr.	CBC Bed % S	FBM Limestone lb/hr Feed Rate	CBC Flue Gas, lb/hr	System* S Release Rate, lb/hr.	% S Capture		Reg. Coal Rate, lb/hr	Reg. Air Rate	Reg. Fly Ash C %	Reg. T, °F	Flue Gas SO$_2$ %	Flue Gas lb/hr Rate	Output S, lb/hr	Reg. Bed S %	Reg. O$_2$ %	CO %	HC ppm	Reg. CO$_2$ %
7/12	1000	N/A	N/A	N/A	N/A	N/A	N/A		N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	1200	15.7	N/A	153	N/A	N/A	N/A		N/A	N/A	N/A	1900	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	1400	19.1	N/A	100	N/A	N/A	N/A		N/A	N/A	N/A	1900	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	1600	17.8	N/A	6	N/A	N/A	(System)	(FBM)	N/A	N/A	N/A	1900	6.3	N/A	N/A	N/A	N/A	0.01	3000	21
	1800	15.1	2.75	64	1216	2.8	81.5	86.3	31	262	N/A	1950	N/A	282	19.5	2.56	N/A	N/A	N/A	N/A
	2000	14.3	3.64	80	1207	2.45	82.9	88.4	28	257	N/A	1950	N/A	276	N/A	2.44	0.8	N/A	N/A	N/A
7/12	2200	14.6	3.6	120	1304	3.65	68.6	92.7	29	257	N/A	1980	5.8	277	17.6	0.60	0.3	0.27	200	25
	0000	16.7	N/A	84	1075	2.6	84.5	93.8	32	261	N/A	2040	7.16	281	22.1	N/A	0.3	N/A	N/A	N/A
7/13	0200	16.8	2.99	15	1154	2.[illegible]	86.3	93.4	30	262	N/A	1950	8.5	282	26.3	1.02	N/A	0.27	200	25
	0400	16.3	2.66	52	N/A	N/A	N/A	N/A	30	N/A	N/A	2040	3.9	N/A	12.	0.91	0.5	N/A	N/A	N/A
	0600	16.8	N/A	50	1030	1.56	90.7	92.5	30	261	N/A	2000	3.9	281	12.	N/A	0.4	0.23	150	25
	0800	15.7	3.0	1	1015	3.85	75.5	78.4	27	241	N/A	2000	7.24	259	20.6	0.44	0.7	0.19	100	25
	1000	14.4	2.86	68	N/A	1.85 FBM	N/A	79	N/A	N/A	N/A	1920	7.1	N/A	20.2	3.35	0.4	N/A	N/A	N/A
	12	15.9	4.16	91	N/A	N/A	N/A	N/A	38	N/A	N/A	1800	N/A	N/A	N/A	2.84	N/A	N/A	N/A	N/A
	14	16.8	N/A	104	N/A	N/A	N/A	N/A	N/A	N/A	N/A	1900	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	16	16.3	N/A	55	1285	2.03	87.5	91.9	36	336	N/A	1940	N/A	360	N/A	N/A	N/A	N/A	100	24
	1800	17.5	N/A	44	N/A	N/A	N/A	N/A	36	N/A	N/A	1940	6.6	N/A	26.1	N/A	1.2	0.19	180	N/A
	2000	18.9	3.78	0	N/A	N/A	N/A	N/A	36	N/A	N/A	1860	6.42	N/A	N/A	2.08	0.7	0.31	380	25
7/13	2200	18.2	4.05	0	1355	2.89	84.1	85.7	38	336	N/A	1930	N/A	361	25.4	1.77	0.7	N/A	N/A	N/A
	2400	14.9	N/A	95	1355	N/A	N/A	N/A	N/A	N/A	N/A	1900	N/A	N/A	N/A	N/A	0.5	N/A	N/A	N/A
7/14	0050	N/A	N/A	0	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	0400	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	
	1800	N/A	N/A	34	N/A	N/A	N/A	N/A	18	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	2010	17.1	N/A	34	N/A	N/A	N/A	N/A	18	N/A	N/A	1970	2.3	N/A	N/A	N/A	5.	N/A	480	23
	2200	17.6	3.56	64	N/A	N/A	N/A	N/A	29	N/A	N/A	1940	5.	N/A	N/A	1.36	N/A	N/A	300	25
	2400	18.3	N/A	63	N/A	N/A	N/A	N/A	38	N/A	N/A	2040	5.5	N/A	N/A	N/A	0.5	0.36	600	25
7/15	0200	17.3	N/A	68	N/A	N/A	N/A	N/A	36	N/A	N/A	1970	N/A	N/A	N/A	N/A	N/A	N/A	350	15
	0400	21.0	N/A	53	N/A	N/A	N/A	N/A	37	N/A	N/A	2010	N/A	N/A	N/A	N/A	N/A	0.04	120	15
	0600	18.0	N/A	53	N/A	N/A	N/A	N/A	37	N/A	N/A	1980	N/A	N/A	N/A	3.04	N/A	0.08	80	N/A
	0800	17.7	2.39	53	1300	2.57 (2.12 FBM)	85.5	88.0	37	N/A	N/A	1980	1.45	N/A	N/A	1.68	1.0	1.04	200	12
	1000	19.7	3.62	98	N/A	N/A	N/A	N/A	37	N/A	N/A	2010	1.8	N/A	N/A	1.01	1.0	0.36	100	15
	1200	21.0	2.88	60	N/A	N/A	N/A	N/A	37	N/A	N/A	2030	2.9	N/A	N/A	0.24	0.5	N/A	N/A	N/A
	1400	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	2020	N/A	N/A	N/A	N/A	0.5	N/A	N/A	N/A
	1435	20.8	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
7/15	1800	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A

* Value is for sum of sulfur in flue gas from FBM and CBC unless noted "FBM"

TABLE 10 (CONTINUED)

		FBM1																CBC1								
Day	Time	T, °F	Bed Level In	Coal Rate lb/hr	Air Rate lb/hr	Fly Ash C %	Flue Gas SO_2 ppm	Flue Gas lb/hr	Bed S %	Bed Ca %	Fly Ash % S	Fly Ash % Ca*	O_2 %	CO %	HC ppm	CO_2 %	Steam lb/hr	O_2 %	Ash C %	SO_2 ppm	Air Rate lb/hr	System C Burnup % Overall	T, °F	CO ppm	CO_2 %	HC ppm
7/15	2000	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A		N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	2200	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A		N/A	N/A	N/A	N/A	N/A	8	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	0000	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A		N/A	N/A	N/A	N/A	N/A	9.7	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
7/16	0200	1750	N/A	N/A	N/A	N/A	210	N/A	N/A	N/A	N/A		1.9	0.18	500	8.	N/A	6	N/A	N/A	N/A	N/A	1940	N/A	N/A	N/A
	0400	1450	N/A	536	N/A	N/A	250	N/A	N/A	N/A	N/A		N/A	N/A	N/A	N/A	N/A	8	N/A	N/A	N/A	N/A	1930	N/A	N/A	N/A
	0600	1640	16	536	7605	N/A	N/A	8008	N/A	N/A	N/A		3.	N/A	N/A	7.5	N/A	6	N/A	N/A	1480	N/A	2030	N/A	N/A	N/A
	0900	1470	N/A	509	N/A	N/A	430	N/A	N/A	N/A	N/A		5.8	0.11	200	8.5	N/A	3	N/A	N/A	N/A	N/A	1900	N/A	N/A	N/A
	1000	1480	N/A	493	N/A	N/A	680	N/A	2.86	36.47	N/A		7.2	N/A	100	7.5	4000	6	N/A	9000	N/A	N/A	2010	7300	13.	40
	1200	1470	N/A	N/A	N/A	N/A	580	N/A	N/A	N/A	N/A		N/A	0.06	N/A	7.5	3100	4.6	N/A	7000	N/A	N/A	1880	22000	12.5	100
	1400	1470	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A		N/A	0.13	N/A	N/A	3650	4	N/A	N/A	N/A	N/A	1970	N/A	N/A	N/A
	1600	1510	N/A	502	N/A	N/A	960	N/A	N/A	N/A	N/A		N/A	0.36	130	8.5	3450	2	N/A	850	N/A	N/A	2000	N/A	7.1	40
	1800	1470	N/A	450	N/A	N/A	710	N/A	N/A	N/A	N/A		N/A	0.09	110	8.6	3500	2	22.2	850	N/A	N/A	1980	N/A	14	40
	2000	1470	N/A	N/A	N/A	N/A	710	N/A	N/A	N/A	N/A		N/A	0.08	150	8.6	3150	3	N/A	240	N/A	N/A	2000	N/A	11.5	40
	2200	1450	17.5	484	7600	N/A	720	7963	2.22	N/A	N/A		N/A	0.19	160	8.	3200	8	N/A	580	N/A	N/A	1800	N/A	N/A	40
	2400	1460	N/A	465	N/A	N/A	N/A	N/A	2.63(2230)		N/A		7.2	N/A	N/A	8.1	3400	10	N/A	320	N/A	N/A	1980	N/A	N/A	N/A
7/17	0200	1470	N/A	520	N/A	N/A	310	N/A	N/A	N/A	N/A		N/A	0.19	160	8	3400	10	N/A	N/A	N/A	N/A	1950	N/A	N/A	40
7/17	0400	1700	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A		N/A	0.18	130	8	4200	10	N/A	N/A	N/A	N/A	2000	100	N/A	35
	0600	1670	N/A	525	N/A	N/A	N/A	N/A	N/A	N/A	N/A		3.2	N/A	900	7	3800	10	N/A	N/A	N/A	N/A	1960	100	N/A	30
	0800	1510	N/A	506	N/A	N/A	500	N/A	1.12	N/A	N/A		9.4	0.05	50	8.6	3000	10	N/A	N/A	N/A	N/A	1860	N/A	N/A	N/A
	1000	1510	N/A	446	N/A	N/A	340	N/A	N/A	N/A	N/A		5.6	0.09	310	8.5	3400	9.8	N/A	N/A	N/A	N/A	1990	N/A	N/A	N/A
	1200	1470	N/A	N/A	N/A	38.[illegible]	580	N/A	1.68	N/A	1.56		N/A	0.08	100	8.5	3250	10	8.5	N/A	N/A	99.	2010	N/A	N/A	N/A
	1400	1490	N/A	507	N/A	N/A	580	N/A	Salt In	N/A	N/A		4.6	0.09	160	9.	3500	9	N/A	N/A	N/A	N/A	2040	N/A	N/A	N/A
	1600	1510	N/A	520	N/A	26.2	300	N/A	2.2	17.25	2.02		6.4	0.27	100	8.5	3500	10	2.7	N/A	N/A	99.	1840	N/A	N/A	N/A
	1800	1520	N/A	N/A	N/A	28.4	310	N/A	2.51	N/A	1.85		7.4	0.23	55	9.6	3500	10	6.5	N/A	N/A	99.	1980	N/A	N/A	N/A
	2000	1440	17	477	7605	31.[illegible]	N/A	7962	1.39	N/A	1.61	15.2	N/A	N/A	N/A	N/A	3300	7	2.6	N/A	N/A	99.	1950	N/A	N/A	N/A
7/17	2200	1525	17.5	472	7150	N/A	490	7504	N/A	N/A	N/A		8.5	0.08	50	10.5	3300	10	N/A	N/A	960	N/A	1960	N/A	N/A	N/A
	2400	1530	N/A	473	N/A	38.[illegible]	490	N/A	1.9	N/A	1.61	17.5	8	0.13	N/A	11.5	3400	10	3.0	N/A	N/A	99.	2030	N/A	N/A	N/A
7/18	0200	1600	N/A	490	5795	N/A	N/A	6162	N/A	N/A	N/A		1.6	N/A	N/A	N/A	3400	N/A	N/A	N/A	1095	N/A	2010	N/A	N/A	N/A
	0400	1470	N/A	445	N/A	38.9 w/ salt	500	N/A	N/A	N/A	N/A		4	N/A	N/A	N/A	3150	N/A	N/A	N/A	N/A	N/A	1970	N/A	N/A	N/A
	0600	1500	N/A	399	6045	36.[illegible]	200	6336	2.20	N/A	2.34	23.88	3	0.35	N/A	N/A	3000	N/A	9.9	N/A	1295	99.	1950	N/A	N/A	N/A
	0800	1450	N/A	366	N/A	N/A	230	N/A	N/A	N/A	N/A		2.2	0.10	70	7.8	2950	N/A	N/A	N/A	N/A	N/A	1960	N/A	N/A	N/A
	1000	1440	16	387	5600	N/A	210	5900	N/A	N/A	N/A		6	0.09	80	7.6	2950	N/A	N/A	N/A	N/A	N/A	1910	N/A	N/A	N/A
	1200	1440	N/A	392	N/A	40.[illegible]	540	N/A	1.82	25.39	1.97		6	0.12	70	8	2950	4.2	7.9	N/A	N/A	99.	1920	N/A	N/A	N/A
	1400	1390	N/A	392	N/A	N/A	560	N/A	N/A	N/A	N/A		N/A	0.09	30	7	2750	3	N/A	N/A	N/A	N/A	1500	N/A	N/A	N/A
	1600	1500	N/A	360	N/A	N/A	N/A	N/A	New Coal	N/A	N/A		N/A	0.09	20	9	N/A	4.0	N/A	N/A	N/A	N/A	1780	N/A	N/A	N/A
	1800	1500	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A		N/A	0.09	50	9	2950	4.4	N/A	N/A	N/A	N/A	1960	N/A	10.8	N/A
	2000	1540	N/A	450	[illegible]	N/A	[illegible]	7662	3.92	N/A	N/A		6.5	0.23		8.8	3100	10.2	N/A	420	890	N/A	2010	300	13	30
7/18	2200	1520	N/A	[illegible]	7235	[illegible]	430	7604	3.92	N/A	1.84		8.3	0.09	120	9	2950	N/A	7.6	630	1165	99.	1885	300	7	40
	2400	1520	14	[illegible]	[illegible]	[illegible]	380	7314	N/A	N/A	2.74	23.9	N/A	0.09	150	9.5	3100	N/A	20.4	630	863	95.1	1910	300	9	30
7/19	0200	1470	N/A	447	N/A	N/A	380	N/A	N/A	N/A	N/A		5.4	0.19	320	9.5	3200	N/A	N/A	N/A	N/A	N/A	1960	N/A	N/A	N/A
	0400	1460	N/A	N/A	N/A	[illegible]	310	N/A	3.36	N/A	2.64		4.	0.13	150	9.5	3250	N/A	4.9	N/A	N/A	99.	1960	N/A	N/A	N/A

*Carbon Free Basis

TABLE 10 (CONTINUED)

Day	Time	System S Input In Coal, Lb/Hr	CBC Bed % S	Limestone Lb/Hr Feed Rate	CBC Flue Gas, Lb/Hr	S Rel. Lb/Hr	Capture %	Reg. Coal Rate Lb/Hr	Reg. Air Lb/Hr.	Reg. % C In Ash	Reg. Temp. °F	Reg. SO_2 %	Reg. Flue Gas Lb/Hr	Output S Lb/Hr	Reg. Bed S %	Reg. O_2 %	Flue Gas: CO %	Flue Gas: HC ppm	CO_2 %
7/15	2000	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	Rate	Reg.	N/A	N/A	N/A	N/A	N/A
	2200	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	0000	N/A	N/A	88	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
7/16	0200	N/A	N/A	88	N/A	N/A	N/A	N/A	N/A	N/A	1620	N/A	N/A	N/A	N/A	1.2	N/A	N/A	N/A
	0400	N/A	N/A	88	N/A	N/A	N/A	N/A	N/A	N/A	1660	0.8	N/A	N/A	N/A	N/A	0.42	350	11
	0600	N/A	N/A	88	1520	2.2 (FBM)	N/A	N/A	280	N/A	2040	0.7*	300	N/A	N/A	1.6	0.87	600	7.5
	0800	18.3	N/A	88	N/A	N/A	N/A	46	N/A	N/A	1980	1.0*	N/A	N/A	N/A	N/A	0.13	300	10
	1000	16.3	0.93	N/A	N/A	N/A	N/A	0	N/A	N/A	1650	0.85	N/A	N/A	0.58	2	N/A	N/A	11
	1200	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	2090	0.92*	N/A	N/A	N/A	0.8	N/A	450	8
	1400	N/A	N/A	N/A	N/A	N/A	N/A	37	N/A	N/A	1940	1.45	N/A	N/A	N/A	2.5	0.07	80	13
	1600	17.9	N/A	64	N/A	N/A	N/A	40	N/A	N/A	2030	1.3	N/A	N/A	N/A	0.4	N/A	140	12.5
	1800	16.1	N/A	56	N/A	N/A	N/A	37	N/A	N/A	1980	1.7	N/A	N/A	N/A	0.2	0.33	N/A	11
	2000	N/A	N/A	71	N/A	N/A	N/A	41	N/A	N/A	1940	1.8	N/A	6.2	N/A	0.9	0.73	450	10.5
	2200	17.1	2.52	71	N/A	6.2 (FBM)	64 (FBM)	33	290	N/A	1940	1.65	310	5.7	N/A	0.4	0.52	370	12.7
	2400	16.6	N/A	64	N/A	N/A	N/A	37	N/A	N/A	1940	1.65	N/A	5.7	N/A	0.2	0.19	170	13
7/17	0200	18.5	N/A	0	N/A	N/A	N/A	40	N/A	N/A	2000	1.45	N/A	N/A	N/A	0.6	0.51	120	11
	0400	N/A	N/A	0	N/A	N/A	N/A	41	N/A	N/A	1900	N/A	N/A	N/A	N/A	0.1	N/A	N/A	N/A
	0600	18.6	N/A	0	N/A	N/A	N/A	39.7	N/A	N/A	1890	N/A	N/A	N/A	N/A	0.6	0.27	N/A	N/A
	0800	18.0	1.19	302	N/A	N/A	N/A	39.7	N/A	N/A	1900	2.1	N/A	N/A	1.32	N/A	N/A	N/A	N/A
	1000	16.1	N/A	N/A	N/A	N/A	N/A	41	N/A	N/A	2000	3.5	N/A	N/A	N/A	1.1	N/A	N/A	15
	1200	N/A	1.49	149	N/A	N/A	N/A	35	N/A	N/A	1960	3.	N/A	N/A	0.31	0.8	1.13	200	23
	1400	18.0	N/A	0	N/A	N/A	N/A	39	N/A	N/A	1970	N/A	N/A	N/A	N/A	0.4	N/A	N/A	N/A
				Salt on															
	1600	18.4	2.0	0	N/A	N/A	N/A	38	N/A	N/A	2050	2.6	N/A	N/A	0.27	1.4	0.52	100	21
	1800	N/A	1.82	63	N/A	N/A	N/A	34	N/A	N/A	1960	2.4	N/A	8.4	1.39	0.8	0.32	250	16
	2000	16.8	1.51	61	N/A	2.7 (FBM)	84 (FBM)	31.5	295	16.8	1930	2.4	315	8.4	0.35	0.3	1.13	650	15
	2200	17.0	N/A	61	1000	4.0 (FBM)	76.5 (FBM)	43	384	12.5	2050	3.	412	13.6	N/A	0.4	0.48	375	22
	2400	16.9	1.66	0	N/A	N/A	N/A	40	N/A	N/A	1860	3.8	N/A	12.1	N/A	N/A	0.56	730	23
7/18	0200	17.4	N/A	0	1135	3.3 (FBM)	81.1 (FBM)	37	265	N/A	1900	2.9	209	9.2	N/A	N/A	0.73	1600	19.5
	0400	15.9	N/A	24	N/A	N/A	N/A	38	N/A	N/A	1860	N/A	N/A	N/A	N/A	N/A	0.52	150	18
	0600	14.1	0.35	0	1335	1.39 (FBM)	90.1 (FBM)	37	265	N/A	1980	N/A	209	N/A	N/A	N/A	0.50	1400	N/A
	0800	13.4	N/A	67	N/A	N/A	N/A	40	N/A	N/A	1880	1.4	N/A	4.5	N/A	N/A	N/A	1200	N/A
	1000	13.8	N/A	67	N/A	1.39 (FBM)	90.2 (FBM)	37	N/A	12.3	1860	1.4	N/A	N/A	N/A	2	0.53	340	12
	1200	14.3	N/A	63	N/A	N/A	N/A	41	N/A	13.6	1900	1.8	N/A	N/A	0.22	1.2	0.87	600	14.5
	1400	N/A	N/A	63	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	0	N/A	N/A	N/A	N/A	N/A	N/A
	1600	N/A	N/A	N/A	N/A	N/A	N/X	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	1800	N/A	N/A	N/A	N/A	N/A	N/A	29	N/A	N/A	1900	2.6	N/A	9.4	N/A	2	N/A	N/A	14.3
	2000	20.8	N/A	69	930	4.0 (FBM)	80.8 (FBM)	33	305	N/A	2070	2.7	326	9.7	1.64	2	0.35	100	16
7/18	2200	17.4	N/A	64	1205	4.0 (FBM)	77 (FBM)	29	305	27.7	2040	2.7	324	9.6	0.56	1.6	0.56	300	17
	2400	21.2	2.97	55	900	2.24 (FBM)	89.5 (FBM)	32	325	36.4	2000	N/A	345	10.2	N/A	1.4	N/A	N/A	N/A
7/19	0200	20.8	N/A	152	N/A	N/A	N/A	35	N/A	N/A	1940	N/A	N/A	N/A	N/A	1.8	N/A	N/A	N/A
	0400	N/A	3.3	103	N/A	N/A	N/A	35	N/A	11.7	1900	4.5	N/A	14.5	1.43	N/A	0.42	1400	20

* Possible Air Infiltration

TABLE 13 (CONTINUED)

		FBM																CBC			System C				
Day	Time	T, °F	Bed Level In.	Coal Rate	Air	Fly Ash Carbon %	SO_2 ppm	Flue Gas lb/hr	Bed S %	Bed Ca %	Fly Ash % S	Fly Ash % Ca	O_2 %	CO %	HC ppm	CO_2 %	Steam lb/hr	O_2	Ash C %	SO_2 ppm	Burnup Overall %	T, °F	CO ppm	CO_2 %	HC ppm
7/19	0600	1460	N/A	N/A	N/A	N/A	220	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	3250	N/A	N/A	N/A	N/A	1920	N/A	N/A	N/A
	0800	1490	N/A	450	6700	N/A	380	7040	N/A	N/A	N/A	N/A	5.5	0.19	500	9.7	3300	4.	N/A	N/A	N/A	1975	N/A	N/A	N/A
	1000	1460	19.5	465	N/A	35.4	320	N/A	3.58	N/A	N/A	N/A	6.	0.25	550	9.6	3200	N/A	18.6	N/A	96.0	1900	N/A	N/A	N/A
	1200	1410	N/A	354	N/A	32.6	280	N/A	4.03	29.4	1.62	N/A	8.	0.23	520	9.6	3050	5.5	13.07	N/A	97.2	1880	N/A	N/A	N/A
	1400	1460	18.5	563	N/A	22.2	N/A	N/A	2.37	N/A	1.48	N/A	7	N/A	500	10.5	3150	6.6	10.7	N/A	98.1	1960	N/A	N/A	N/A
	1600	1400	N/A	483	N/A	N/A	420	N/A	N/A	N/A	1.09	10.6	8	0.22	N/A	10	2850	7.6	N/A	500		1430	1600	11.	220
	1753	N/A	15	483	N/A	N/A	N/A	N/A	5.02	N/A	N/A	N/A	6	0.16	200	12	2900	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A

*With 12% ash coal

TABLE 13 (CONCLUDED)

														FLUE GAS:			
Day	Time	System S Input	CBC Bed % S	FBM Limestone lb/hr. Feed Rate	CBS Flue Gas lbs/hr	FBM S Release lb/hr	S Capture %	Reg. Coal Lb/Hr	Reg Air Flow lbs/hr	Reg. Fly Ash C%	Reg. T, °F	Reg. SO_2 %	Reg. Output S Lb/Hr	Reg. Bed S %	Reg. O_2 %	Reg. CO %	Reg. HC ppm	Reg. CO_2 %
7/19	0600	N/A	N/A	112	N/A	N/A	N/A	30.4	N/A	N/A	1890	3.6	N/A	N/A	N/A	0.45	900	21.5
	0800	20.8	N/A	62	N/A	2.94 (FBM)	86.3 (FBM)	31	N/A	N/A	1895	3.6	N/A	N/A	N/A	0.03	1000	22
	1000	21.4	3.38	62	N/A	N/A	N/A	30	N/A	28.5	1880	N/A	8.8	0.36 (1100)	N/A	N/A	N/A	N/A
	1200	16.6	2.60	61	N/A	N/A	N/A	30.5	N/A	25.6	2050	2.5	N/A	0.83 (1300)	N/A	0.12	250	18
	1400	25.7	2.18	0	N/A	2.2 (FBM)	91.3 (FBM)	33	N/A	N/A	1940	3.69	N/A	N/A	0.5	1.12	N/A	21
	1600	22.2	N/A	0	N/A	N/A	N/A	31	N/A	37.3	1870	1	N/A	N/A	N/A	1.8	900	18
	1753	N/A	N/A	0	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	0.86	N/A	N/A	N/A	13

(18 determinations). The other coal (used later in the test) was "Rivesville" (see Section 5.6). With the Powhatan coal, average regenerator fly ash carbon content was 13.4% (5 determinations). Based on this result a CBC treatment of the Regenerator output ash may not be necessary. During the first 38 hours of the test, the CBC ash collector was faulty. The result of this on the burnup measurements is that they are questionable.

Average REG bed sulfur content was 1.11% (26 determinations). Average FBM bed sulfur content was 3.05% (33 determinations). The REG SO_2 output and bed S contents imply a bed circulation rate of about 670 lb/hr (about 0.6 FBM bed changes per hour), which is consistent with the transfer screw system operating rate plus natural circulation between fluidized beds. In a boiler system initially designed to incorporate both a CBC and a regenerator, the REG-FBM slot lengths would be less than in the present system, and natural circulation would be enhanced. The present Regenerator is an add-on device. The need for screw system operation (an expedient) would be reduced.

Median SO_2 emission from the FBM was 320 ppm with the 3.3% S coal, equivalent to a capture of 89% in the FBM, based on coal input to the system. Since this capture is less than that experienced in Run 168H, it appeared possible that a system operating change had acted to reduce capture. One possibility was that the REG fly ash (which contains an average of 4% S and up to 5.4% S and was fed to the CBC), also contained a poison which was recaptured in the CBC under oxidizing conditions. To flush possible poisons from the system, an unnaturally high limestone feed rate was imposed on the system (averaging 55 lb/hr) and bed material periodically withdrawn and discarded.

To facilitate carbon balance calculations for various portions of this test, fly ash streams were analyzed periodically for carbon, calcium, and sulfur. These results are collected in Table 11. These materials were not analyzed for arsenic. For coal No. 1 (the 6.1% ash, 3.2% S Powhatan coal) the average S/Ca ratio in the FBM fly ash was 0.15 (7 determinations). The median Ca content of these ashes was 10.5% (22 determinations). For coal No. 2 (the 12.4% ash, 3.85% S Rivesville coal) the FBM fly ash Ca contents were all below 8.4% in line with the heavier inert loading, and the average S/Ca ratio in these ashes was 0.24 (6 determinations). The fly ash from the CBC tended to have a S/Ca ratio of 0.08 regardless of coal origin (12 determinations).

The dilution of the $CaSO_4$ in the fly ash by elutriated CaO in the CBC is probably due to the relatively high gas velocities used. Dissociation of the $CaSO_4$ content of the fly ash or bed in the CBC would be undesirable, from an air pollution standpoint.

At one point in the test, salt addition to the FBM was performed. Salt was screw fed from a weighing hopper and mixed with the coal feed. The result was an immediate 48% reduction in SO_2 emissions and a 32% reduction in fly ash carbon content (see Table 10, entry "7/17-1400").

TABLE 11

FLY ASH CHARACTERISTICS - RUN 171H

Vessel	Time	Input* %C	%Ca (as rec'd)	%S	Vessel	Output** %C	%C	%S
FBM	12-1850***	49.6	15.6	1.71	CBC	N/A	37.65	(1705)
	12-1905	N/A	N/A	N/A	CBC	N/A	N/A	2.4
FBM	12-2050	29.7	29.47	1.97	CBC	0	N/A	N/A
	12-2200	N/A	N/A	N/A	CBC	1.	42.2	2.41
FBM	12-2300	36.5	16.2	2.18	N/A	N/A	N/A	N/A
FBM	13-0210	53.5	10.1	1.64	CBC	1.	32.2	2.68
	13-0456	N/A	N/A	N/A	CBC	1.	N/A	2.63
FBM	13-0835	49.9	6.9	1.4	CBC	1.	33.5	2.32
FBM	13-1045	36.2	15.1	1.37	CBC	1.8	25.75	1.72
FBM	13-1250	27.5	10.05	2.29	CBC	0.7	39.31	1.8
FBM	13-1800	N/A	11.4	N/A	CBC	N/A	20.3	N/A
FBM	13-2030	21.8	10.05	1.23	CBC	0.51	28.8	(2045)
FBM	13-2230	27.3	10.05	1.09	CBC	1.	35.6	2.56
FBM	14-0712	41.9	9.68	1.68	N/A	N/A	N/A	N/A
FBM	14-2330	24.55	11.7	1.0	CBC	N/A	17.3	1.82
FBM	14-2400	24.6	N/A	N/A	CBC	3.9 (2300)	27.7	2.2
FBM	15-0845	50.4	10.6	N/A	CBC	11.95	21.6	1.55
FBM	15-1045	50.6	9.1	1.76	CBC	2.6	36.9	3.43 3.20
FBM	15-1245	46.8	9.4	3.14	CBC	3.9	35.6	2.28 2.30
	16-1700	N/A	N/A	N/A	CBC	22.2	20.	0.83
FBM	17-1130	38.8	9.7	1.56	CBC	8.5	22.5	N/A
FBM	17-1535	26.2	16.95	2.02	CBC	2.7	23.5	N/A
FBM	17-1830	28.4	15.5	1.85	CBC	6.5	22.1	2.2
FBM	17-1830	N/A	11.4	N/A	CBC	N/A	20.3	N/A
FBM	17-2030	3.12	10.45	1.61	CBC	2.6	26.5	1.24
FBM	17-2330	38.2	10.85	1.61	N/A	N/A	N/A	N/A

TABLE 11 (continued)

		Input				Output		
Vessel	Time	%C	%Ca (as rec'd)	%S	Vessel	%C	%C	%S
FBM	17-2345	N/A	N/A	N/A	CBC	3.	19.8	1.42
FBM	18-0400	38.9	NLA	N/A	N/A	N/A	N/A	N/A
FBM	18-0515	36.8	15.1 w/salt	2.34	CBC	9.9	29.1	2.36
FBM	18-1145	40.1	NLA	1.97	CBC	7.9	NLA	2.72
FBM	18-2115	38.6 new coal	6.97	1.84	CBC	7.6	17.8	2.0
FBM	18-2300	N/A	15.6	2.74	CBC	20.4	9.	1.54
FBM	18-2400	34.9	N/A	N/A	N/A	N/A	N/A	N/A
FBM	19-0435	33.4	14.5	2.69	CBC	4.9	21.31	1.99
FBM	19-1100	N/A	6.01	1.62	CBC	N/A	26.92	1.93
FBM	19-1000	35.4	N/A	N/A	CBC	18.6	N/A	N/A
FBM	19-1300	N/A	6.18	1.48	CBC	N/A	13.05	1.20
FBM	19-1200	32.6	N/A	N/A	CBC	13.07	N/A	N/A
FBM	19-1500	22.2	8.3	1.09	CBC	10.07	12.64	1.55

* Input to CBC ** Output of CBC

*** "12-1850" means July 12, 1972, @ 1850 hours

Note: NLA = Sample expended, no longer available

Regenerator fly ashes were also analyzed for calcium (average 23%, 10 determinations and carbon, median 14% 11 determinations). Fly ash calcium values were converted to $CaCO_3$ equivalent output from each reactor (see Table 12).

There is a slight dependence of FBM fly ash calcium output on limestone feed rate to the FBM. This may be attributed to a certain fraction of the makeup limestone particles being inherently weak. Dead slow on the limestone feeder represented 60 lb/hr, with the particular drive gearing in use. Rates below this value reflect intermittent operation of the feeder. When limestone feed rate is less than 25 lb/hr, median calcium fly ash output is calculated to be 18 lb/hr $CaCO_3$ equivalent. When feed rate is above 52 lb/hr, median calcium fly ash output is calculated to be 26 lb/hr $CaCO_3$ equivalent. The CBC calcium output tends to be higher than the FBM, due to its higher superficial velocity* as set in this test.

The FBM fly ash median S content was about 1.7%. This ash could not be fed to a CBC operating with an agglomerating ash bed** at temperatures above 2,000°F without a substantial excess of lime present, as SO_2 emissions up to 2,400 ppm might result. However, it is possible that the functions of carbon burnup, ash agglomeration and bed regeneration can be recombined in one vessel, and development of this concept is needed.

The system heat balance data are collected in Table 13. With the particular configuration of add-on regenerator in use in this test, radiation losses are high.

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* See Table 12 for typical velocity values.

** Reference Goldberger Patent, U.S. #3,171,369.

TABLE 12

CALCIUM BALANCE, FBM/CBC/REG TEST NO. 171-H

Day	Time	$CaCO_3$ Fed lb/hr	Output in FBM Fly Ash $CaCO_3$ equiv., lb/hr	Output in CBC Fly Ash $CaCO_3$ equiv., lb/hr
12	1800	64	41	55
12	2000	80	75	70
12	2300	84	33	46
13	0200	16	26	46
13	0800	1	13	48
13	1000	68	23	26
13	2000	0	12	41
13	2200	0	14	64
15	0800	53	25	28
15	1000	98	24	75
15	1200	60	27	73
17	1800	63	27	30
17	2000	61	15	32
17	2330	0	18	
18	0600	12±12	24	37
18	2000	69	19	35
18	2400	55	60	19
19	1000	62	16	92
19	1400	0	22	67
Median		60	24	46
Average		44	27	48
Typical superficial vel., ft/sec			9	15

TABLE 13 Heat Balance Data Run 171-H

DAY TIME	12 1800	12 2000	12 2400	13 0200	13 0845	13 1000	13 2000	13 2200	15 0800	15 1000	15 1200	17 1800
FBM Bed temp., °F	1525	1480	1500	1540	1540	1480	1460	1480	1500	1460	1500	1520
SO_2 capture %	86	88	94	93	78	79	91	86	88	88	88	84
Carbon Burnup %	99+	99+	99+	99+	99+	99+	99+	99+	98.7	99+	99+	99+
Reg SO_2 %	6.3	5.8	7.1	8.5	3.9	7.1	6.4	6.4	1.5	1.8	2.9	N/A
Total preheated air, lb/hr	5545	5300	5340	5420	5880	5880	6675	7295	8000	8000	8000	N/A
KBtu/hr	411	367	360	376	452	452	463	523	574	574	574	N/A
CBC Air, lb/hr	1130	1120	985	1065	925	925	1195	1265	1300	1300	1300	N/A
FBM coal heating value, KBtu/hr	5577	5315	6246	6299	5918	5367	6220	6758	6561	7349	7873	N/A
Total input, KBtu/hr	6345	5987	6970	7056	6724	6120	7155	7780	7579	8331	8926	N/A
Steam Absorption, KBtu/hr	3045	3147	3147	3248	3248	3147	3350	3350	3198	3451	3451	N/A
Circ. H_2O Abs. KBtu/hr	1437	1416*	1422	1451	1422	1422*	1440	1501	1517	1517	1517*	3552
Total (H_2O + St.)	4482	4563	4569	4699	4670	4569	4790	4851	4715	4968	4968	N/A
% of coal value	80	85	73	74	78	85	76	71	71	67	63	N/A
Flue gas lcss, KBtu/hr	707	664	652	662	672	672*	852	866	882	889	891	N/A
Blowdown allowance	15	16	16	17	17	16	17	17	17	17	17	N/A
Hydrocarbon loss	60	60	60	60	100	112	122	122	64	64	17	18
Limestone calcining	50	62	66	13	N/A	53	N/A	N/A	42	77	213†	N/A
CO loss, KBtu/hr	25	45	47	46	83	107	118	118	61	61	47	N/A
CBC carbon loss	48*	45*	54*	54*	51*	45*	54*	58*	92	62*	122	N/A
% of coal value	0.8	0.8	0.8	0.8	0.8	0.8	0.7	0.8	1.3	0.8	67*	58*
Hot fly ash loss	16	16	14	14	13	13	12	16	15	20	0.8	N/A
FBM dust coll. carbon loss	71	41	41	74	51	30	15	22	64	73	21	N/A
Lift air loss	16	19	N/A	N/A	N/A	N/A	27	21	21	21	73	28
Radiation and unaccounted	494	151	1147	1054	615	104	685	1166	1074	1582	21	N/A
Total losses	1452	1057	2031	1981	1602	1099	1902	2406	1290	2789	1959	N/A

*Estimated

† Coal Feeder
50% plugged

6.4.3 Arsenic Analyses

In test 171H, an apparent loss in sorbent activity was observed. When fly ash from a regenerator is fed to a Carbon Burnup Cell, along with the primary boiler cell fly ash, an interference with SO_2 capture in the primary boiler cell occurs. This interference may be due to adding the REG to the system. We hypothesize the REG fly ash may contain a poison. In a test to be described in the next section of this report, this hypothesis was tested by discarding the REG fly ash and using an unpoisoned bed. The result was gratifying, up to 97% SO_2 capture.

A possible poison is arsenic, which is present in coal to the extent of 0.01% or less (mostly as arsenopyrites, FeAsS) and would be captured in the FBM as calcium arsenate. The REG flue gas when cooled would then contain a smoke of As_2O_3, free or absorbed on carbon.

A series of arsenic analyses was therefore performed. See Appendix Q. Condensed results are:

	µg/g
Coal (Powhatan)	7
Limestone	0
FBM bed (average of 12 determinations increasing with time)	8.4
REG output fly ash (median of 7 determinations)	10.9
REG bed (median of 13 determinations)	7.6
CBC bed (2 determinations)	4.4

The colorimetric analytical procedure used is given in Reference 6. Good results were obtained with all materials except coal, where excessive foaming usually occurred. A reasonable correlation existed between REG bed As and S contents.

The results are consistent with a bed poisoning mechanism in which As in coal is captured by lime in the FBM as calcium arsenate which hypothetically interferes with SO_2 capture (as a shell?); desorbs in the regenerator under high temperature - low O_2 conditions and is carried in REG fly ash as arsenous oxide (as a smoke or absorbed on carbon) and is recaptured in the CBC under oxidizing conditions. Apparently, once arsenated, the bed is "regenerated" (dearsenated) only with difficulty*. It is significant that the FBM bed As content increases with time. The definitive experiment which needs to be performed is to perturb the capture of SO_2 by lime in a batch experiment using the existing equipment for aqueous additive solutions. Sodium arsenate could be added, to see if limestone activity is reduced. The effect of salt addition on As removal from the beds should also be researched.

6.5 FBM/REG Test (Test No. 172H)

Following completion of the long duration FBM/CBC REG Test (No. 171-H), it was desired to test system operation with the REG fly ash, together with whatever poisons it contained, discarded. An 8-hour test was therefore performed, in which the FBM and REG only were used. A fresh -8+16 mesh limestone bed was used. Fly ash was discarded. Four FBM fly ash samples analyzed 25.6, 23.2, 15.9 and 23.4% carbon. Regenerator flue gas SO_2 content was 10% volume. FBM flue gas SO_2 varied between 70 and 110 ppm. FBM CO_2, O_2, NO, CO and temperature ran in the usual range. These results are consistent with facile capture of SO_2 by regenerated lime in the absence of a bed poison. The 70 ppm SO_2 value corresponds to 97% capture.

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* It is reasonable that at a given temperature, "DEA" (As_2O_3 mobility) would proceed less rapidly than "DES" (SO_2 mobility, see page 6-15) from molecular diffusivity and molecular weight considerations.

6.6 Bed Particle Size Determinations

Bed material samples were withdrawn at regular intervals during the FBC and FBM/CBC tests for calcium and sulfur analyses as well as sieve analysis. Carbon analysis showed zero carbon in the bed, as would be expected. The six sieve fractions from the Run C-321 samples are tabulated in Appendix Table G-1 and shown graphically in Figure G-1. No significant size change is indicated by the raw data or the probability plot. When treated statistically, the weight mean particle size defined by

$$\bar{d} = \Sigma\, \bar{d}_i x_i$$

(where x_i = weight fraction of total in i^{th} sieve fraction.)

shows a small size reduction as run duration increases. (See Appendix Figure G-2). On a trial basis, the data were also treated according to

$$\bar{d}' = \frac{1}{\sum \frac{x_i}{d_i}}$$

but the data scatter is greater by this method due to the anomalous -25 mesh fraction at 3.7 hours sampling time. A reliable determination of the effect of salt on attrition rate is not possible from this test.

The Run C-322 samples were analyzed similarly (see Appendix Table G-2 and Figures G-3 and G-4). At the higher combustion intensity and coal and air rates an apparent attrition rate occurs, where $\dot{d}$ is the time rate of diameter decrease:

Test No.	Air Rate lb/hr	Coal Rate lbs/hr	Bed Depth Inches	$\dot{d}$, μm/hr	Fly Ash Rate lb/hr
321	610	41	16	2.2	2.9*
322	670	64	22	9	25.

* exclusive of salt adding

Characteristics of the five sieve fractions for the Run C-323 samples are tabulated in Appendix Table G-3. There is no significant size change between the 2.7, 3.7 and 4.7 hour samples. "5.1 hr." is an aged sample. The 1.7 hr. sample has a much smaller mean diameter, reason not known.

The Run C-324 samples were analyzed similarly (see Appendix Table G-4 and Figure G-5). In the 2.1, 3.1 and 4.1 hour samples, particle mean diameter increases at the rate of 28 μm/hr, presumably due to coal ash absorption. These are periods of low fly ash generation. Following salt addition, the particle mean diameter decreases at the rate of 6 μm/hr, and the average fly ash generation rate is higher. The 6.96 hour diameter is anomalous.

The FBM Run B-18 samples were analyzed similarly (see Appendix Table G-5). Particle mean diameter decreases at an apparent average rate of 14 μm/hr. The data scatter is unsatisfactory in the 9.4 and 10 hour samples.

Selected bed particle size data from Run 168H are collected in Appendix G, Table G-6, and Figures G-6 and G-7. Since these beds are all at least 36 hours old*, mean particle sizes are considerably less than the batch FBC beds (C series) (nonregeneration operation) and FBM B-18 beds described earlier. Elutriation rate of fines depends on air rate and particle density (sulfur and contaminants content) among other factors. Mean particle size depends upon makeup limestone rate (-8+20 mesh) and elutriation rate, as well as rate of fines production which may be related to severity of regeneration as well as coal particle size.

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* i.e., the test has been in progress for at least this time before the sample in question.

Selected Run 171H beds were particle size analyzed and the results are listed in Appendix G, Table G-7. The mean particle size fell during the first six days. The diameter percentage loss average per day was 1.7 over the first five days. During the 7th and 8th days Rivesville coal was fed; it contains 1/4" top size rocks and these affect the bed size consist.

7. PRELIMINARY FLUIDIZED-BED BOILER DESIGNS, HEAT BALANCES, AND COST ESTIMATES.

7.1 Preliminary 30 Megawatt Boiler Designs, Heat Balances and Cost Estimates

The following design outlines are based upon (1) the incremental boiler cost, including dust collector with SO_2 abatement system and (2) the overall power plant cost, for a multiple module fluidized-bed boiler installation vs. a conventional pulverized coal installation.

7.1.1 30 Megawatt Packaged Boiler Concepts (Figure 26, 27 and 28)

The 30 megawatt designs, for 300,000 lb steam/hr, 1,270 psig, 925°F have been developed in three successive stages, based essentially on proven pilot boiler (FBM) operation.

(a) RV-I, Figure 26: A non-regenerative limestone (once thru) unit with four primary boiler cells and a Carbon Burnup Cell, utilizing primary superheater elements in the boiler section with secondary superheater elements in the carbon burnup section. Coal and limestone feed are mixed via overhead pressurized "run-around" mass flow conveyor to metering screws which, in turn, deliver the coal/limestone mixture to vertical feeders extending through the boiler convection section to inbed split 180° coal feed sections.

(b) RV-II, Figure 27: A 3-combustion zone unit (boiler, limestone regenerator, Carbon Burnup Cell) with primary superheater split between freeboard and inbed immersed sections* of primary cells. An attemperator

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* Freeboard is also commonly termed "slop zone" in fluidized-bed boilers.

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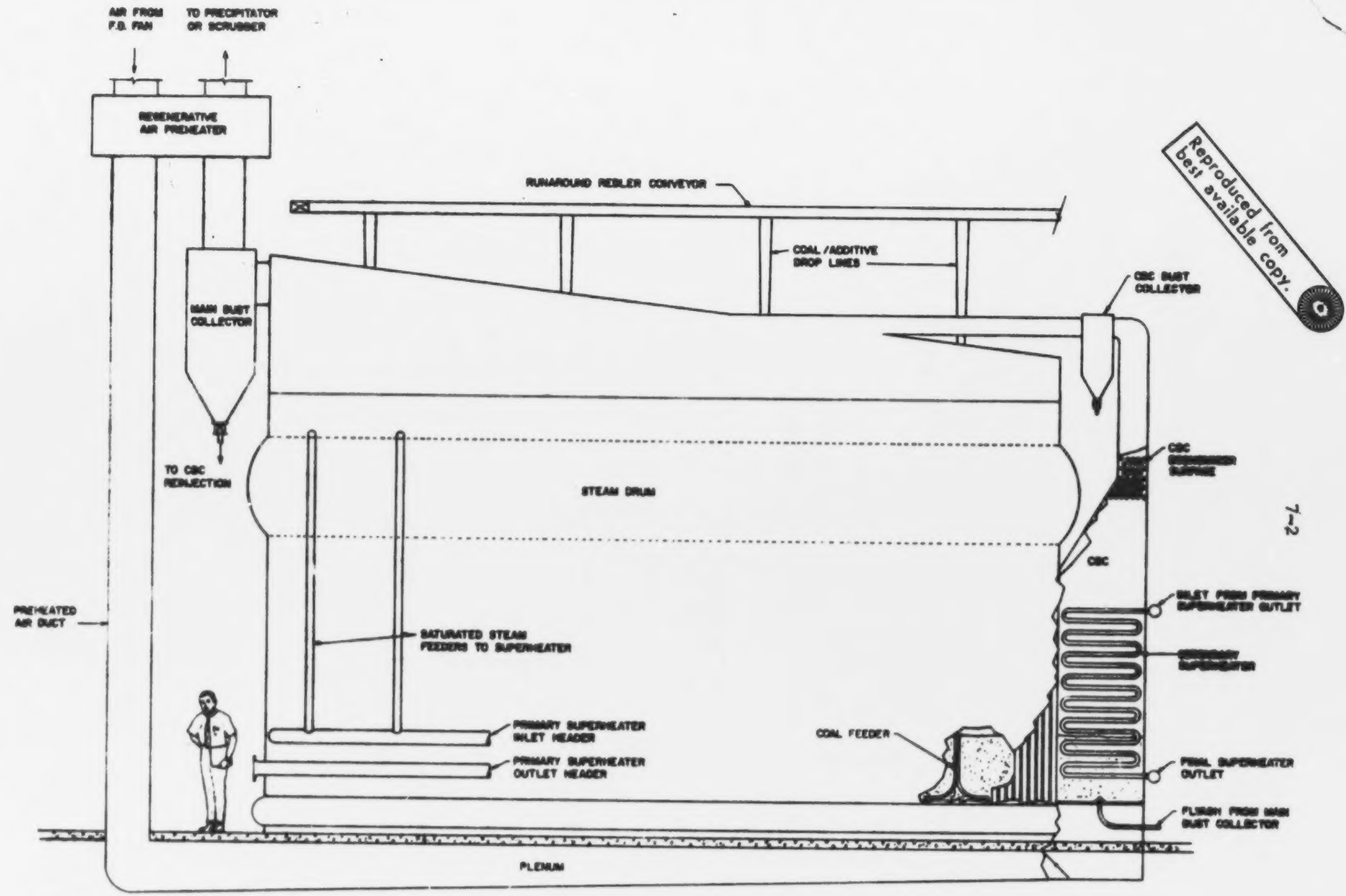
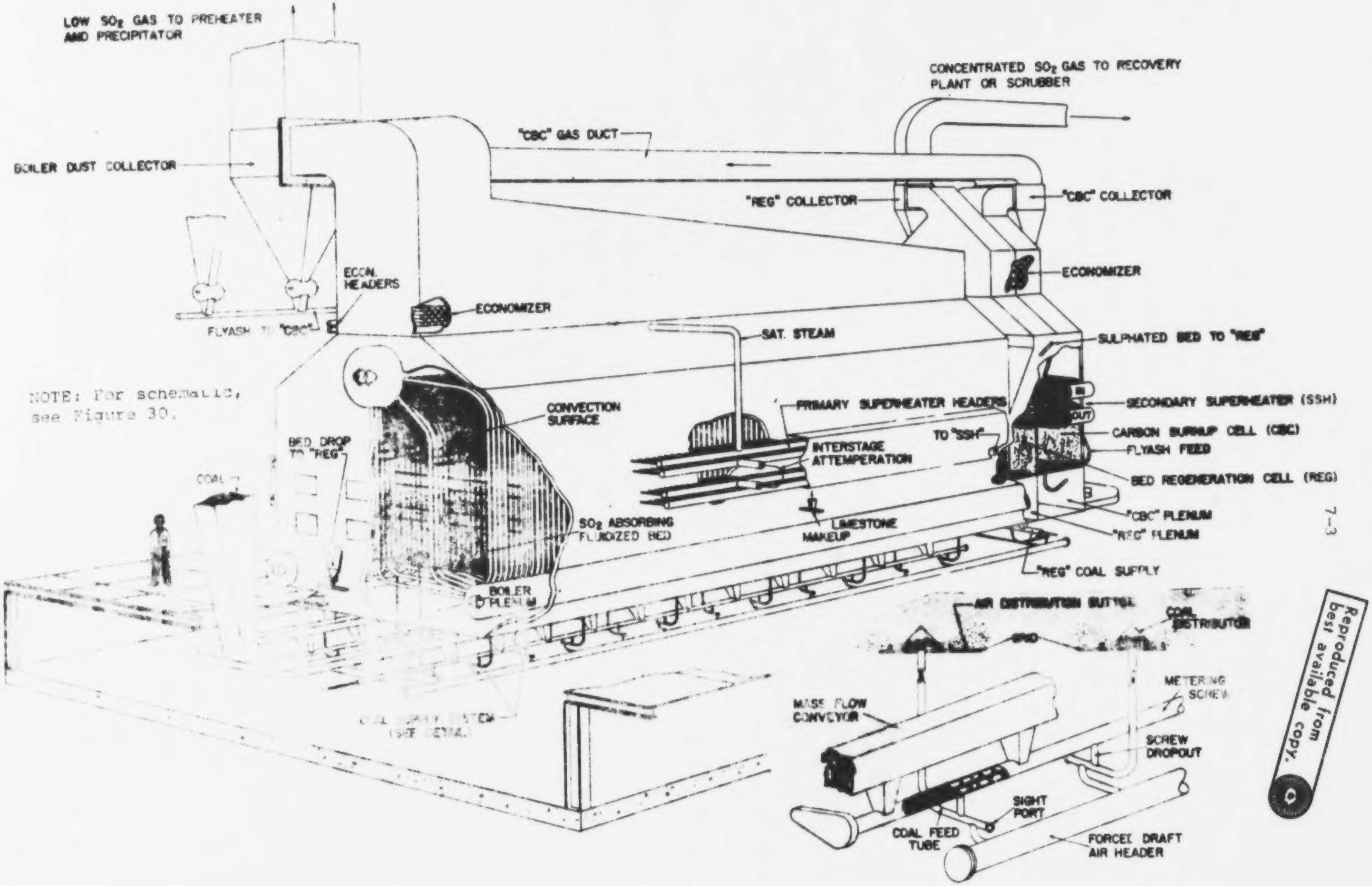

AIR FROM F.D. FAN
TO PRECIPITATOR OR SCRUBBER
REGENERATIVE AIR PREHEATER
RUNAROUND REBLER CONVEYOR
COAL/ADDITIVE DROP LINES
CBC DUST COLLECTOR
MAIN DUST COLLECTOR
TO CBC REINJECTION
STEAM DRUM
CBC SURFACE
CBC
PREHEATED AIR DUCT
SATURATED STEAM FEEDERS TO SUPERHEATER
INLET FROM PRIMARY SUPERHEATER OUTLET
SECONDARY SUPERHEATER
PRIMARY SUPERHEATER INLET HEADER
PRIMARY SUPERHEATER OUTLET HEADER
COAL FEEDER
FINAL SUPERHEATER OUTLET
FLUSH FROM MAIN DUST COLLECTOR
PLENUM

FIGURE 26. COAL-FIRED FLUIDIZED-BED UTILITY BOILER, FACTORY ASSEMBLED, 300,000 LB/HR, 1270 PSIG, 925°FTT, RV-I.


LOW SO2 GAS TO PREHEATER AND PRECIPITATOR
BOILER DUST COLLECTOR
"CBC" GAS DUCT
CONCENTRATED SO2 GAS TO RECOVERY PLANT OR SCRUBBER
"REG" COLLECTOR
"CBC" COLLECTOR
ECONOMIZER
ECON. HEADERS
ECONOMIZER
FLYASH TO "CBC"
SAT. STEAM
SULPHATED BED TO "REG"
IN
OUT
SECONDARY SUPERHEATER (SSH)
PRIMARY SUPERHEATER HEADERS
CONVECTION SURFACE
TO "SSH"
CARBON BURNUP CELL (CBC)
INTERSTAGE ATTEMPERATION
FLYASH FEED
BED DROP TO "REG"
BED REGENERATION CELL (REG)
COAL
"CBC" PLENUM
SO2 ABSORBING FLUIDIZED BED
LIMESTONE MAKEUP
"REG" PLENUM
"REG" COAL SUPPLY
BOILER PLENUM
AIR DISTRIBUTION BUTTON
GRID
COAL DISTRIBUTOR
METERING SCREW
MASS FLOW CONVEYOR
COAL SUPPLY SYSTEM (SEE DETAIL)
SCREW DROPOUT
SIGHT PORT
COAL FEED TUBE
FORCED DRAFT AIR HEADER

NOTE: For schematic, see Figure 30.

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FIGURE 27. 300,000 LB/HR PACKAGED FLUIDIZED-BED BOILER (1270 PSIG, 925°F) FOR HIGH-SULFUR COAL, RV-II.

is provided between the slop zone and inbed portion of the primary superheater for superheater control. Final superheat is provided within the carbon burnup and regenerator sections. Economizer surface is provided in a separate package installed over the boiler drum. Express gas flow is provided for even above bed draft and for optimum jetted bed particle return.

Coal feed is of the through-grid mushroom distributor type. Coal is supplied by a single-point-supplied run-around pressurized mass flow conveyor, from which two (one on each side) metering screws receive and deliver a controlled quantity of coal to each mushroom feeder, as shown in the detail of Figure 27. Limestone feed may be via separate injection point, as shown, or may be mixed in with the coal entering the mass flow conveyor. Air distribution is by means of 15° downthrow nozzle buttons, as used in the experimental equipment, which preclude sifting of bed material back into the plenum upon shutdown.

The limestone and flue gas cycle is described in the schematic detail of Figure 27, as demonstrated in the FBM pilot boiler. Low SO_2 flue gas from the 1500-1600°F boiler beds pass through the slop zone, convection banks and economizer section into the reinjection collector and thence to a regenerative air preheater and fly ash precipitator. Flue gas from the Carbon-Burnup Cell passes through the discard fly ash collector to the same air preheater and precipitator. Concentrated SO_2 flue gas, in typically about 4% of total flue gas stream, is directed, via a dust collector, to the sulfuric acid recovery plant or to a small lime scrubber.

The low temperature boiler beds (absorption beds) pick up sulfur primarily in the form of $CaSO_4$ in the calcined lime particles. Boiler (absorption bed) material is continuously removed from the front of the unit and transported to the regenerator section operating at 1900-2000°F and about 0.5% O_2. Regenerated limestone (as CaO) with some residual sulfur content passes by gravity back into the low temperature boiler bed via slots in the boiler/regenerator tube wall, as demonstrated in pilot FBM boiler. The Carbon-Burnup Cell, operating at 2000°F, 3% O_2, "floats" on the regenerator with minimal amount of bed material exchange (as necessary to equalize level).

Fuel supply to the regenerator is by separately controlled coal feed supply. Fuel for the Carbon-Burnup-Cell is the fly ash from the reinjection collector. Lightoff is by single burner, "burrowing" into the bed of any coal containing combustion zone, followed by automatic propagation as demonstrated in the FBM pilot boiler. Lightoff for this concept will be in the regenerator section, since the regenerator communicates with all four boiler cells and the Carbon-Burnup Cell, thus obviating the need for more than one burner.

A predicted heat balance for this concept is presented by Table 14.

(d) RV-III, Figure 28: A modification of the RV-II design with the following modifications:

(1) Superheater tubes are in the boiler section, consisting of primary convection surface followed, via an attemperator, by the secondary superheater immersed in the fluid bed.

TABLE 14

RV-II HEAT BALANCE AND SURFACE SUMMARY

Circuit	Duty MBtu/hr		Temperature $^\circ$F	Surface ft^2
Secondary SH	22		2000 ↔ 2000 792 → 925	425
Primary SH	61		1550 ↔ 1550 581 → 792	2650
Inbed Evap.	162		1550 ↔ 1550 581 → 581	3570
Convection Evap.	37		1550 → 1108 581 ↔ 581	4250
Economizer	48	(Over boiler)	1108 → 730 525 ← 385	5850
		(Over CBC/REG.)	2000 → 730 525 ← 385	1720

A 78.5% effective Ljungstrom preheater lowers 341,000 lb/hr of boiler and CBC flue gas from 730°F to 275°F to raise 328,000 lb/hr of air from 70°F to 470°F.

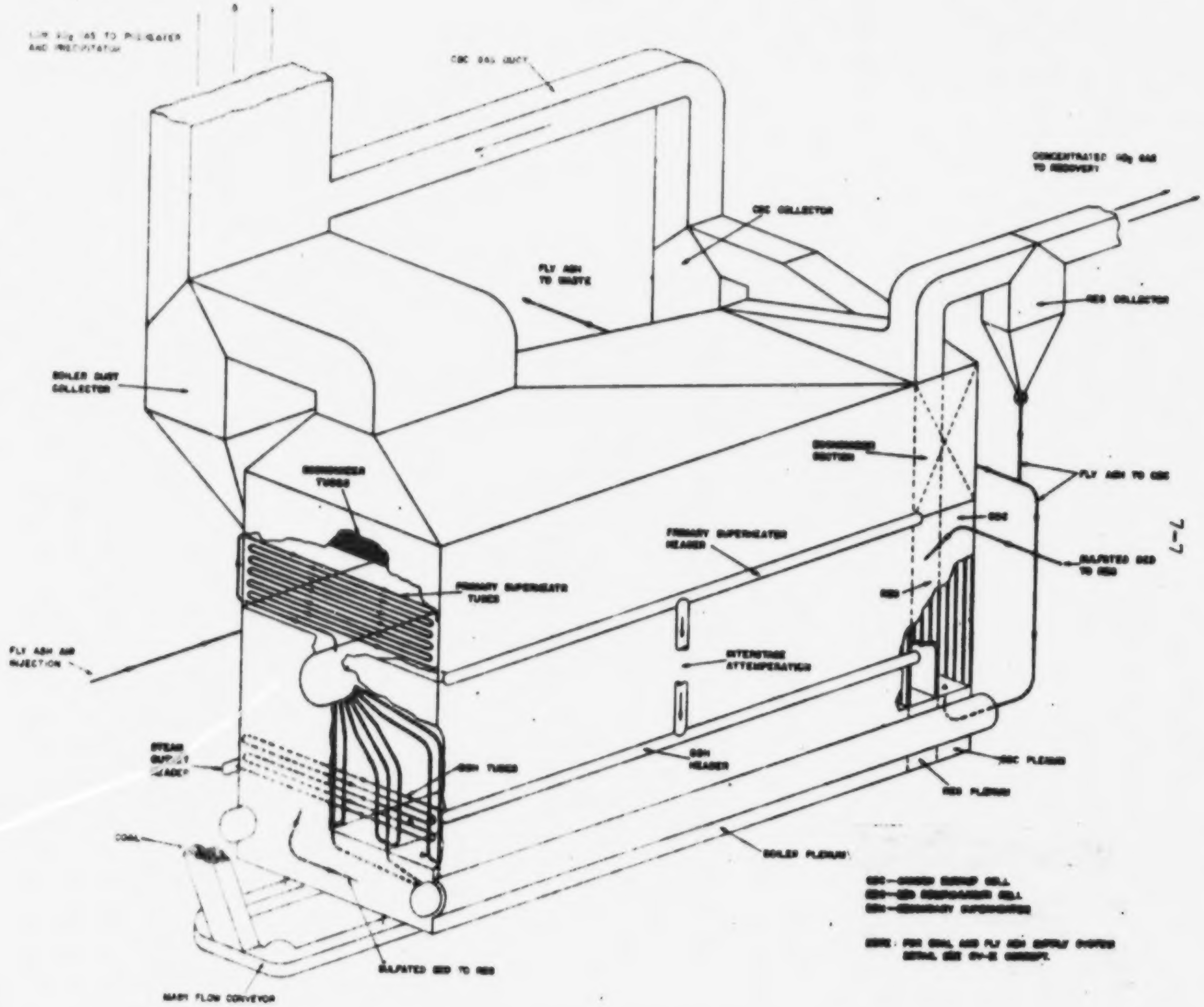
LOW NO_x GAS TO PURIFIER AND PRECIPITATOR
CBC GAS DUCT
FLY ASH TO WASTE
CBC COLLECTOR
CONCENTRATED NO_x GAS TO RECOVERY
RBD COLLECTOR
BOILER DUST COLLECTOR
ECONOMIZER TUBES
ECONOMIZER SECTION
FLY ASH TO CBC
CBC
PRIMARY SUPERHEATER HEADER
SULFATED BED TO RBD
PRIMARY SUPERHEATER TUBES
RBD
FLY ASH AIR INJECTION
INTERSTAGE ATTEMPERATION
STEAM OUTLET HEADER
SSH TUBES
SSH HEADER
CBC PLENUM
RBD PLENUM
COAL
BOILER PLENUM
SULFATED BED TO RBD
MASS FLOW CONVEYOR

FIGURE 28. 300,000 LB/HR PACKAGED FLUIDIZED-BED BOILER, 1270 PSIG, 925°F FOR HIGH-SULFUR COAL, RV-III.

(2) The closely spaced boiler convection section is eliminated (drum ligament problems);

(3) Combustion zone heat pickup is augmented by two additional open spaced tube rows which do not interfere with maintenance accessibility, plus additional superheater surface;

(4) Extended economizer banks provide for cooling REG and CBC off gases;

(5) Elimination of expensive alloys in secondary superheater.

Predicted heat balance for the RV-III unit is outlined in Table 15.

7.1.2 30 Megawatt Boiler Cost Estimates, Fluidized-Bed (RV-III vs. Pulverized Coal)

The incremental cost* of the packaged RV-III fluidized-bed boiler is summarized by Table 16. Item breakdowns are presented by Appendix L.

The incremental cost of a field erected pulverized coal unit is estimated, per Appendix M, at $2,538,000 plus $1,000,000 for a wet limestone SO_2 abatement system.

7.1.3 Overall 30 Megawatt Power Plant Cost Estimates, Fluid Bed vs. Pulverized Coal

A comparison of the estimated overall 30 Megawatt Power Plant cost, is provided in Table 17.

7.2 Preliminary 300 Megawatt Fluidized Boiler Concept, Heat Balances and Cost Estimates

The unit presented herein as Figure 29 is an 8-boiler module unit with integral limestone regeneration and carbon burnup section. Capacity is 1,900,000 lbs/hr @

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* The incremental cost excludes items common to any boiler system of equal capacity. Feed pumps are an example.

TABLE 15

RV-III HEAT BALANCE AND SURFACE SUMMARY

Section	Duty $\overline{\text{M}}$Btu	Temperature Diag °F			Surface Sq. Ft.
Boiler	190	1550 587	↔ ↔	1550 587	6,550
Secondary Superheater	57	1550 658	↔ →	1550 925	2,070
Primary Superheater	35	1550 694	→ ←	1177 587	8,300
		TOTAL STEAM GENERATION			16,900 ft^2
Heat Traps[1]					
"REG" Economizer	2	2000 393	→ ←	730 385	330
CBC Economizer	12	2000 421	→ ←	715 393	2,320
Boiler Economizer	41	1177 525	→ ←	715 429	16,700
Ljungstrom	43	715 604	→ ←	250[2] 70	66,100 ft^2

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[1] Actual total economizer duty is 46 $\overline{\text{M}}$Btu versus 55 $\overline{\text{M}}$Btu as listed. The 9 $\overline{\text{M}}$Btu difference is a "safety factor" which may be applied to manufacturer's margin and unaccounted. Theoretical surface is on the order of 16,900 ft^2 vs. the 19,300 ft^2 surface as tabulated.

[2] Allowing for air leakage in the Ljungstrom, corrected exit gas temperature becomes 220°F.

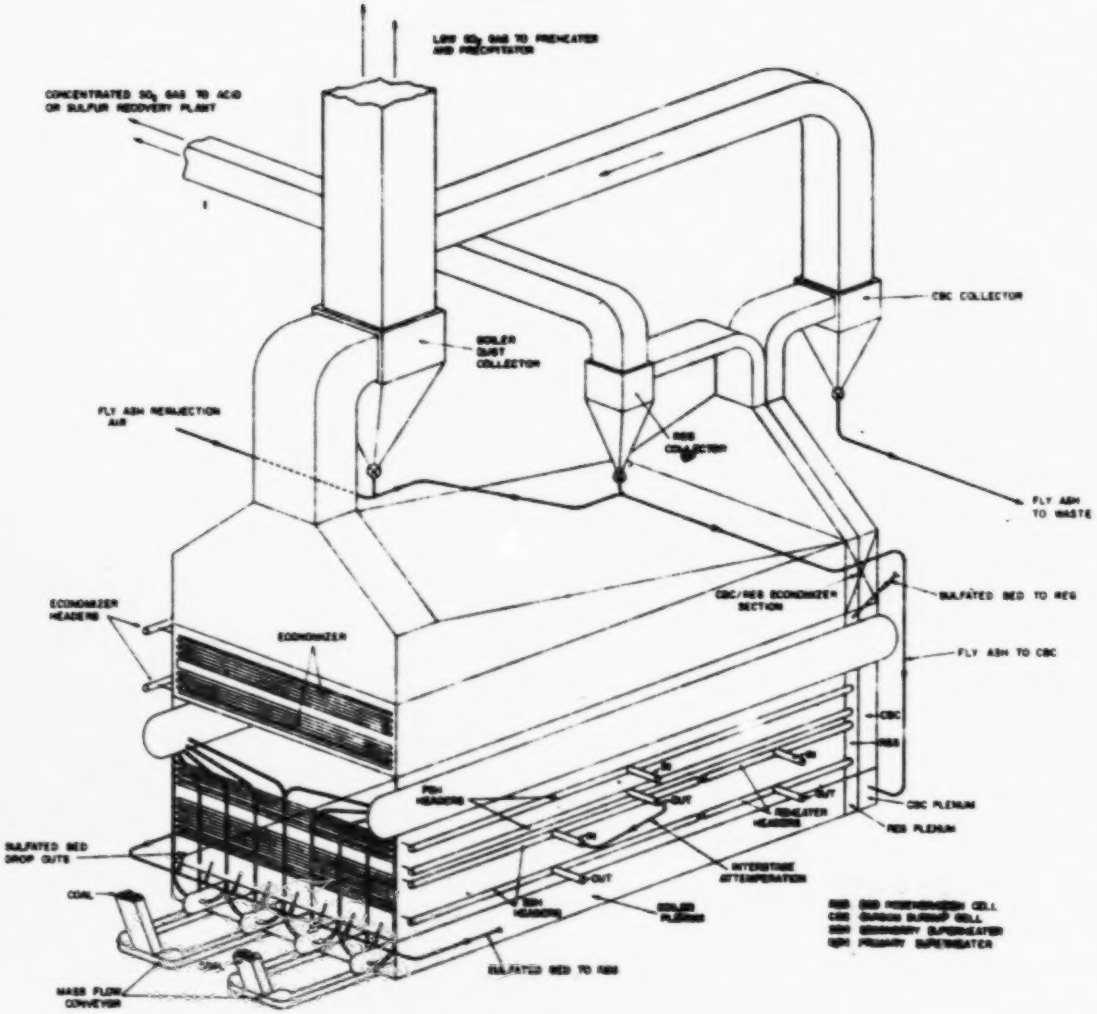

LOW SO2 GAS TO PREHEATER AND PRECIPITATOR
CONCENTRATED SO2 GAS TO ACID OR SULFUR RECOVERY PLANT
BOILER DUST COLLECTOR
CBC COLLECTOR
FLY ASH REINJECTION AIR
REG COLLECTOR
FLY ASH TO WASTE
CBC/REG ECONOMIZER SECTION
SULFATED BED TO REG
ECONOMIZER HEADERS
ECONOMIZER
FLY ASH TO CBC
CBC
REG
IN
OUT
PSH HEADERS
REHEATER HEADERS
CBC PLENUM
REG PLENUM
SULFATED BED DROP OUTS
INTERSTAGE ATTEMPERATION
COAL
SSH HEADERS
BOILER PLENUM
REG BED REGENERATION CELL
CBC CARBON BURNUP CELL
SSH SECONDARY SUPERHEATER
PSH PRIMARY SUPERHEATER
SULFATED BED TO REG
MASS FLOW CONVEYOR

FIGURE 29. 300-MW UTILITY BOILER CONCEPT, 1,900,000 LB/HR 2400 PSIG, 1000°F, REHEAT 1,600,000 LB/HR, 650 PSIG, 1000°F.

TABLE 16

INCREMENTAL COST, 300,000 LB/HR
FLUIDIZED-BED BOILER INSTALLATION

BASIC BOILER		COST (Dollars)	WEIGHT (Tons)
1.	Main boiler package (including REG, CBC furnaces and secondary superheater, with insulation, casing, and grid	310,830	165
2.	Primary superheater	44,000	43
3.	Economizer (all sections)	109,950	110
4.	Air plenum, casing enclosure, ducts, etc.	16,930	24
5.	Structural supports, platforms, etc.	13,930	50
6.	Boiler trim	46,703	3
7.	F.D. fan and 700 HP motor drive	21,200	6
	Basic boiler subtotal	563,000	401
	Erection	170,000	
	TOTAL BASIC BOILER	733,000	
SUPPLEMENTAL BOILER PLANT ITEMS, INSTALLED			
8.	Dust collector and precipitator	201,000	
9.	Bed moving system	[illegible],000	
10.	Coal and limestone supply (from bunkers)	65,700	
11.	Lightoff system	3,000	
12.	Ash moving	4.000	
13.	Controls and instrumentation	70,000	

TABLE 16

(Continued)

14.	Air preheater	139,000
15.	Incremental misc. boiler connections, piping	20,000
16.	Incremental boiler connections, electrical	10,000
	Auxiliary boiler plant subtotal	517,000
	Total Boiler Plant (excluding building)	$1,250,000
	Boiler Plant Building	105,000
	TOTAL BOILER PLANT	$1,355,000

Note: This figure excludes constant cost items; e.g., the electric generation plant, common-to-any system auxiliary equipment, land acquisition, etc.

TABLE 17

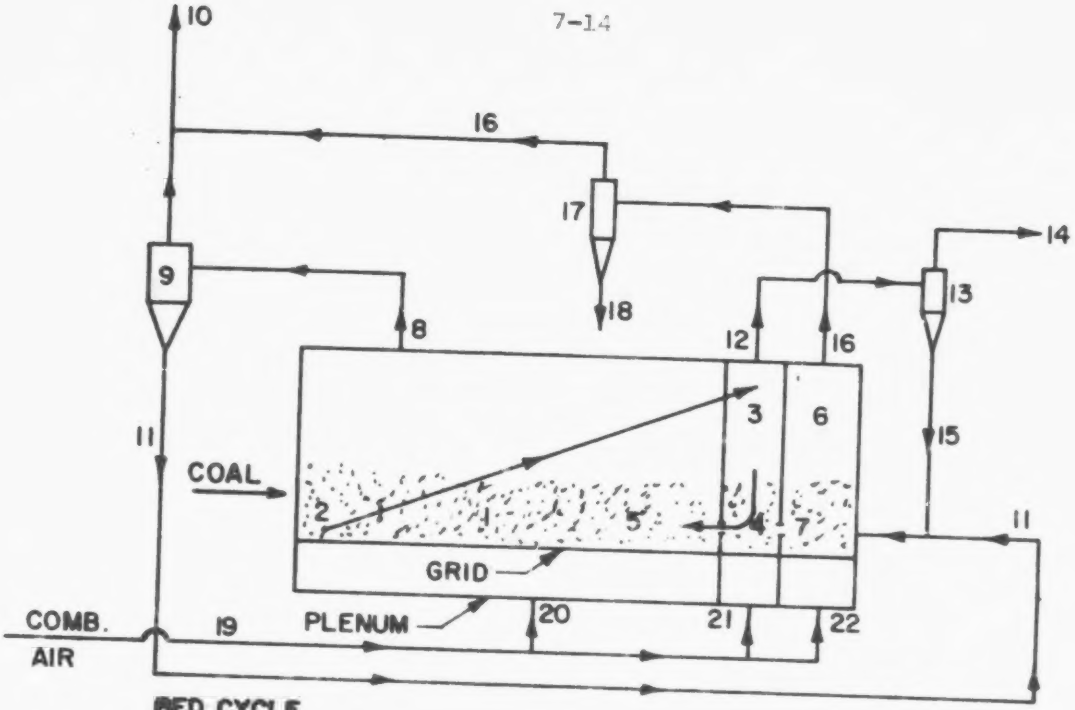
30 MEGAWATT* POWER PLANT COST ESTIMATES

		Costs in Thousand Dollars	
FPC Cate-gory	Items	Packaged F.B. with Integral SO_2 Abatement	Field Erected P.C. with External SO_2 Abatement
310	Land and land rights	50	70
311	Structures and improvements	850	1100
312	Boiler Plant equipment:		
	120 Boiler and accessories	1,044)	2538**
	121 Draft equipment	30)	
	122 Feedwater equipment	220	220
	123 Fuel handling and storage	300	300
	124 Fuel burning equipment	70	Included above**
	125 Ash handling equipment	120	120
	126 Water supply and treating	100	100
	128 Boiler Instruments and Controls	100	Included above**
	129 Boiler plant piping	300	300
	Subtotal of 312: (Boiler Plant Equipment)	2,284	3,578
314	Turbine Generator	2,500	2,500
315	Accessories and electrical equipment	700	700
316	Miscellaneous plant equipment	100	100
	External SO_2 Scrubbing***	350	1,000
	Other expenses	800	800
	TOTAL PLANT COST	7,634	9,848
	Cost per KW Plant Capacity	$ 254	$ 328

* Net plant capacity after subtracting auxiliary loads.

**FPC items 124 and 128 are included in total items 120 and 121 figures.

***Recent quote from Chemical Construction Co.



BED CYCLE

BOILER LIMESTONE BED (1) (CaO) ABSORBS SULFUR AT LOW (1500-1600°F) TEMPERATURE UNDER OXIDIZING CONDITIONS, ACQUIRING A SULFATE ($CaSO_4$) SHELL. PARTIALLY SULFATED BED IS MOVED FROM AREA (2) TO REGENERATOR BED (3) AT HIGH (2000°F) TEMPERATURE, AND LOW OXYGEN CONCENTRATION WHERE $CaSO_4$ REVERTS TO CaO. REGENERATED BED RETURNS TO BOILER BED (1) VIA SLOTS (4). MAKEUP LIMESTONE IS FED AT (5). CBC BED (6) OPERATES INDEPENDENTLY OF OTHER BEDS, EXCEPT FOR RESTRICTED OPENING (7) FOR BED LEVEL EQUALIZATION.

GAS AND FLY ASH CYCLE

BOILER FLUE GAS (8), PASSES TO MAIN COLLECTOR (9), MIXES WITH CBC FLUE GAS (16), THEN TO PREHEATER AND PRECIPITATOR AT (10). CARBON BEARING FLY ASH IS DISCHARGED AT (11) FOR TRANSPORT TO CBC (6). REGENERATOR GAS (12) PASSES TO REG COLLECTOR (13) AND TO SO_2 RECOVERY (A SCRUBBER) AT (14). FLY ASH IS DISCHARGED AT (15) FOR INJECTION TO CBC ALONG WITH MAIN FLY ASH STREAM (11).

CARBON BURNUP CELL GAS (16) PASSES TO CBC COLLECTOR (17) TO MAIN BOILER FLUE GAS DISCHARGE (10). FLY ASH (18) IS REMOVED TO STORAGE SILO.

PREHEATED COMBUSTION AIR DUCT (19) FEEDS PLENUM SECTIONS AT (20), (21) AND (22).

FIGURE 30. GAS, FLY ASH, AND BED REGENERATION SCHEMATIC.

1,000°F followed by 1,600,000 lb/hr of reheat from 650° to 1,000°F. The boiler is approximately 32 ft. wide by 80 ft. long and 30 ft. high (top of economizer).

The gas, fly ash and bed regeneration cycle is schematically indicated and described by Figure 30.

Coal is supplied to two pressurized Redler distributors (see Figure 29) which delivers the coal to four metering screw arrangements which are periodically supplied and discharged to mushroom feeders (see detail in Figure 27. Limestone feed may be via separate inputs into each cell or may be mixed with coal before the Redler distributor. The Redler distributor ("Runaround Redler Conveyor") is designed to move about double the amount of coal required in order to assure positive uniform feed to the metering screws.

Fly ash from the boiler and regenerator dust collectors is pneumatically introduced to the CBC section through the rear wall of the boiler. Experience has indicated that fly ash feed ports may not be located less than 2 ft. apart for uniform distribution. Fly ash from the CBC collector is discharged to silo.

The intent is to discharge the sulfated bed material from the front of each boiler cell and pneumatically transport the same to the regenerator in similar manner as practiced on the pilot boiler (FBM). The regenerated limestone gravitates back to the 8 boiler beds. The CBC merely "floats" on the system; interconnection between CBC and REG is solely for bed level equalization and lightoff purposes.

Evaporation is handled by the natural circulation vertical wall tubes and inclined boiler tubes connecting to the steam drum. At the 2,400 psig conditions, with the superheat and reheat indicated, evaporation (steam generation) is a minor part of the total heat absorption.

Superheat is handled with primary and secondary sections with attemperation between the two sections. Reheat is shown as a single imbedded section; detailed design might indicate a need for a split reheater with attemperature.

Economizer surface is distributed above the steam drum and lowers the gas temperature entering the air heater into the 800°F range.

The heat balance and surface requirements are summarized by Table 18.

Cost estimates for the unit are the subject of Table 19 as broken down in Appendix N. These are the incremental costs for the boiler, controls, coal supply and boiler auxiliaries, only, i.e., the estimated cost for installation in an existing plant. The overall power plant cost estimate is the subject of Table 20. The incremental boiler installation (based on 1971 costs) is estimated at $24 per KW capacity with the overall plant estimated at $124 per KW capacity.

Based upon an annual consumption of 800,000 tons of 4.5% sulfur coal, approximately 64,000 tons of concentrated SO_2 should be delivered to the acid plant for a by-product of 106,000 tons of 93% H_2SO_4.

The concept as shown is based upon an extension of the 30 Megawatt concept described, heat balanced and estimated in Table 14, which, in turn, was based upon

TABLE 18

300 MEGAWATT FLUIDIZED BED BOILER SINGLE BED LEVEL, NATURAL CIRCULATION HEAT BALANCE AND SURFACE REQUIREMENTS

Circuit	Duty, MBtu	Heat Transfer, °F	Surface Sq. Ft.
INBED AND SLOP ZONE			
Main Boiler Evap	532	1550 ↔ 1550 674 ↔ 674	16,000
Reg Evap	29	2000 ↔ 2000 674 ↔ 674	1,200
CBC Evap.	136	2000 ↔ 2000 674 ↔ 674	1,410
Primary Superheater	114	1550 ↔ 1500 694 ← 674	3,480
Secondary Superheater	538	1550 ↔ 1550 760 → 1000	21,400
Reheat	315	1550 ↔ 1550 650 → 1000	11,650
CONVECTION			
Evaporator	40	1550 → 1230 674 ↔ 674	9,365
Overbed Primary SH	195	1550 → 1242 760 ↔ 694 AV	14,750
Main Economizer	273	1178 → 794 620 → 523 AV	91,490
Reg Economizer	12	2000 → 794 523 → 517 AV	2,430
CBC Economizer	75	2000 → 794 517 ↔ 480 AV	14,150
Preheater (Ljungstrom)	343	794 → 234 693 ← 80	454,400

TABLE 19

INCREMENTAL COST OF 1,900,000 LB/HR, 1000°F SINGLE LEVEL, NATURAL CIRCULATION FLUIDIZED-BED BOILER INSTALLATION WITH 1,600,000 LB/HR REHEAT FROM 650°F TO 1,000°F

		Installed Cost
1.	Boiler, complete, including superheater, atemperator, reheater, economizer, grid, nozzles, insulation, trim, and boiler supports	$4,385,000
2.	Instruments and controls	400,000
3.	F.D. fan and drives	86,000
4.	Dust collector and precipitator	825,000
5.	Bed transport	40,000
6.	Coal supply	274,000
7.	Lightoff system	12,000
8.	Fly ash transport	38,000
9.	Air preheater	800,000
10.	Duct work	150,000
11.	Miscellaneous direct boiler piping and electrical	200,000
	TOTAL	$7,210,000

Note: This Table excludes constant cost items, e.g., the electric generation plant, common-to-any system auxiliary equipment, land acquisition, etc.

TABLE 20

OVERALL 300 MW FLUIDIZED BED POWER PLANT CAPITAL COST ESTIMATE

FPC CATEGORY	ITEM		COST IN THOUSAND DOLLARS
310	Land and land rights		290
311	Structures and improvements		4,000
312	Boiler plant equipment:		
120	Boiler and accessories	$6,288	
121	Draft equipment	236	
122	Feedwater equipment	1,050	
123	Fuel handling and storage	2,350	
124	Fuel burning equipment	286	
125	Ash handling equipment	350	
126	Water supply and treatment	136	
128	Boiler instrumentation/control	400	
129	Boiler plant piping	1,800	
	TOTAL ITEM 312		12,896
314	Turbine generator		10,750
315	Accessory electrical equipment		2,430
316	Miscellaneous plant equipment		530
	Other expenses		6,150
	TOTAL PLANT COST		37,046

pilot boiler (FBM) experience. As such, it is two "generations" away from an actual operating unit. Experience with a smaller commercial prototype will undoubtedly change many of the concepts (Figure 29) and design features.

8. SULFUR RECOVERY

Many processes have been suggested for removal or recovery of sulfur oxides from gas streams. For example, about 40 recovery processes were reviewed recently by Arthur G. McKee and Co. (Reference 8). Considering economics and the current state of technology, the use of limestone wet scrubbing appears to offer one of the most dependable emission control systems at this point. The concept is under continuing study and earlier work has been comprehensively reported by Tennessee Valley Authority (Reference 9). The process is thus suitable either for consideration for installation, or as a base case for comparison with other methods, to be applied to the So_2 Acceptor Process Regenerator output gas stream.

TVA has reported (Reference 10) investment estimates for two sizes of generating plants (200 and 1000 MW), for both limestone injection in the boiler followed by wet scrubbing, and for limestone addition at the scrubber. Annual operating cost estimates are also given. These data make possible a rational application of the TVA figures to other sizes. The present study used the TVA relations for the most part. For the fluidized-bed cases, a straight prorating is not suitable, for while the SO_2 load is nearly as large (90%) as the SO_2 load in the corresponding pulverized fuel case, the volume of gas to be treated is relatively very small (4%). Therefore, the investment items were classified as to whether they would be primarily influenced by SO_2 load or by gas load, and special consideration was given to those items governed by gas load.

The chemicals and utility costs were recalculated for the specific conditions of the cases in the present study. Investment costs were escalated to 1971 values using the Machinery and Equipment Index of the Bureau of Labor Statistics.

The estimates of capital and operating costs developed for the limestone treating processes are summarized in Table 21. The cases considered are: Case 1, Fluid bed boiler, 300,000 lb/hr steam, limestone addition at scrubber. Regenerator gas at 450°F and 5.1% SO_2 enters a gas cooler, then enters the scrubber at 175°F. Countercurrent circulating oil is heated from 130 to 420°F. The limestone supply is prepared and mixed with water, and then enters the delay and mixing tank. Slurry contacts the gas and is sulfated, and returns to the surge tank. Some of the slurry is withdrawn from the surge tank and discarded. Scrubbed gas at 85°F is then reheated by the circulating oil to 390°F, and blown by the I.D. fan to the stack or to the boiler plenum.*

Case 2, same size but pulverized fuel boiler, with limestone addition to the boiler. Case 3, fluidized bed boiler as in case 1 but 1,900,000 lb/hr steam. Case 4, same size but pulverized fired boiler.

The basis used for amounts of limestone to be employed were those selected by TVA for the processes. It will be noted that the amounts were carefully reviewed by TVA on the basis of experimental evidence. Particularly in the scrubber addition system, considerable variation was found in performance with the several sources of calcium carbonate. Accordingly, it is to be expected that the limestone sources proposed for commercial use will be evaluated for satisfactory performance. With

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* The reheating is probably not needed for the fluid-bed boiler since the wet gas could be combined with the bulk of gas, still hot, before the stack.

TABLE 21
CAPITAL & OPERATING COSTS
LIMESTONE TREATING PROCESSES
75% CAPACITY FACTOR

Case	1	2	3	4
Type of Plant Addition and Size	Fluid-Bed Scrubber	Pulv. Fuel Boiler	Fluid-Bed Scrubber	Pulv. Fuel Boiler
lb/hr, Steam	300,000	300,000	1,900,000	1,900,000
T Coal, Annual	88,200	88,200	597,000	597,000
Est. Capital Cost Before Credit, $	677,000*	871,000	2,246,000*	3,112,000
After Precipitator Credit, $	677,000	753,000	2,246,000	2,615,000
Annual Operating Costs Before Credits $/yr	321,500	376,800	931,500	1,201,800
$/T Coal	3.65	4.28	1.56	2.09
After Precipitator Credits, $/yr	321,500	346,000	931,500	1,068,800
$/T Coal	3.65	3.92	1.56	1.79

* Includes reheaters which are probably not necessary.

this precaution, it is anticipated that adequate sulfur dioxide removal will be obtained. The scrubber-addition system is used with the fluidized-bed boiler and the small volume of treated gas can be recycled to the boiler plenum for further exposure to the limestone bed at little cost, instead of being vented directly. Flue gas recycle has been shown to have the additional benefit of reducing NO emission from the boiler. In cases 1 and 3, with wet scrubbing - limestone addition to the scrubber, used boiler lime may optionally be added to the delay tank, for additional SO_2 absorption efficiency.

The limestone method of SO_2 recovery secures only part of the fluidized-bed benefit, namely, the small volume of treater gas, but does not fully capitalize on the high SO_2 concentration available at the regenerator exit. Other possible recovery systems would use the concentration advantage as well:

a. Chemico MgO process. Possibly for cases 1-2

b. Sulfur dioxide recovery. (Allied DMA absorption or H_2) absorption). Possibly case 1.

c. Contact H_2SO_4. Possibly case 3.

d. Elemental sulfur. Possibly case 3.

The most logical strategy for handling SO_2 emissions, if there were a free choice, would be conversion to H_2SO_4. The use of limestone or similar processes to eliminate the SO_2 stack emission obtains relief from the air pollution, but it consumes raw materials and yields a waste product to be discarded somewhere. However, the SO_2 is obviously a potential substitute, in most parts of the U.S., for new sulfur. Since most of the sulfur consumed is in the form of H_2SO_4, the broad marketing opportunities should be best for sulfur in this form.

However, in certain places in the U.S. where no local H_2SO_4 market exists, processes may be needed to convert flue gas SO_2 to liquid SO_2 or elemental sulfur which can be transported to markets elsewhere more economically than H_2SO_4.

There are, of course, a number of circumstances which would influence the final choice of process. Cost is the natural criterion, but other factors enter into the choice. A power plant operator may not wish to become involved with sophisticated chemical technology, with different financing or accounting, and with marketing of by products. From a public relations standpoint, there may be reluctance to handle, store or ship materials having the potential for causing corrosion, a public hazard, or environmental damage in case of mishap.

The four boiler cases considered in this study are sufficiently different that different treatments must be considered. Cases 1 and 2 with about 25 tons per day average of H_2SO_4 potential are probably too small for elaborate operations to produce a bulk commodity, although the high concentration of SO_2 in a small volume in Case 1 is favorable. Cases 3 and 4 with about 175 tons per average day of H_2SO_4 potential offer more promise of favorable economics due to scale-up savings, especially with the rich gas in Case 3 employing the fluidized bed.

Water scrubbing was considered as a potential method for concentrating SO_2 when starting at the 3% SO_2 level. There was concern over the sizes of absorbent stream, the size of exchanger and cooler, and the amount of stripping steam indicated (about 4.5% of boiler capability). Because of budget limitations on the contract program, this approach was not followed up further, nor was the alternate dimethylaniline (Allied) scrubbing approach.

An attempt was made to develop design criteria for a contact H_2SO_4 plant for concentrated SO_2 recovery to be applied to (1) a 30 megawatt and (2) a 300 MW limestone regenerating boiler. Based upon a 90% load factor, burning 13,500 Btu/lb coal, with 4.5% S content with 92% SO_2 absorption in the boiler beds, the 30 MW unit will deliver about 8,500 tons of SO_2 to the acid plant which will product about 14,000 tons of 93% H_2SO_4 as a product. Tail gases from the acid plant would be recycled through the boiler (absorbing) beds by means of plenum reinjection. Budget limitations on the contract program prevented completion of a finished contact acid plant design.

Appendix O contains prorating data, capital cost and operating cost estimates for the limestone scrubbing systems described above.

Appendix P contains additional information on the SO_2 recovery processes considered.

9. REFERENCES

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APPENDIX A

FBC SPECIFICATIONS

APPENDIX A. FBC SPECIFICATIONS

1. Air Supply

   Two centrifugal fans in series for 300 cfm at 30" w.g. connected to a smooth 4" diameter conduit 20' long. Air flow is controlled with a gate valve and monitored with a venturi, pitot pressure, static pressure, and temperature measurements.

2. Plenum

   Mild steel, ¼" thickness, 21"x18"x12" outside dimensions with 8" diameter air inlet.

3. Water Column

   Mild steel, ¼" thickness, 24"x20"x36" outside dimensions with 16"x12"x36" inside dimensions.

   A. Wall on inlet air side contains:
      a) One nominal 3" diameter pipe for lightoff burner.
      b) One nominal 1" diameter instrument port.

   B. Left wall (facing air inlet) contains:
      a) One nominal 2" diameter pipe with valve for removal of bed material.
      b) Eight nominal 1" diameter instrument ports at various levels.
      c) One nominal 1" diameter water outlet.
      d) One nominal 2" diameter pressure relief port.

   C. Right wall (facing air inlet) contains:
      a) One rectangular 2"x1" coal feed port.
      b) One nominal 3/4" diameter cooling water inlet.

D. Wall opposite the air inlet contains:

Three nominal 1½" diameter ports.

E. Insulating liner consisting of ¼" ASTM 446 stainless steel extending from the grate to connection with hood (37-3/4"). Internal dimensions are 9-3/8" x 13¼". Kaowool insulation is placed in the nominal 1" annular space.

4. Air Distribution Grid

The grid contains 130 stainless steel air distribution buttons spaced on 1¼" centers, each containing eight drilled ports, 1.089" diameter. The air is discharged downward at an angle of 15° to the horizontal. With the insulating liner installed, 42 of the air distribution buttons are not in service.

5. Water-cooled Hood

The hood is a truncated pyramid 24" x 20" at the bottom and 17" x 17" at the top, with a height of 24" and a flue opening 12" diameter. Material is #10 gauge mild steel. One 4" diameter observation port is provided with 1" diameter water ports and a 2" diameter pressure relief port.

6. Flue System

From the FBC-1 hood, the flue system is run in 12" diameter #10 gauge steel pipe to the induced draft fan. From the fan the pipe is continued at 6" diameter, again #10 gauge steel. All connecting sections are welded. The induced draft fan may be bypassed.

7. Dust Collector

The collector contains two 8" diameter centrifugal collector units with a dust hopper, rotary feeder, and a valve for fly ash removal.

LOCATION OF THERMOCOUPLES FOR FBC

NO.	LOCATION
1.	Forced draft fan air
2.	Plenum air
3.	Bed 1½"
4.	Bed 3"
5.	Bed 7"
6.	Bed 11"
7.	Bed 15"
8.	Flue gas in hood
9.	Air distribution grid
10.	Dust collector outlet
11.	Gas sample line inlet
12.	Flue gas exit
13.	Cooling water in
14.	Cooling water out-hood
15.	Cooling water out-water walls
16.	Isokinetic probe (above lab)
17.	Sample gas discharge
18.	NO_x sampling line
19.	Flue gas exit (over lab)

APPENDIX B

FBM SPECIFICATIONS

APPENDIX B. FBM SPECIFICATIONS

1. Air Supply

One centrifugal fan at 2500 cfm at 50" w.g. connected to air preheater and 12-inch square duct which expands to full width of plenum at inlet. Air is controlled by means of a damper and monitored by an orifice.

2. Plenum

Mild steel, ¼" thickness, 72" x 20¼" x 12" inside inside dimensions with a 6' x 1' air inlet.

3. Boiler Construction

a. Single 20" steam drum
b. Dual 6" lower headers
c. 2½" risers on 4" centers for side walls
d. 4" downcomers (external)
e. 5'4" distance from grid to uninsulated bottom of steam drum
f. Combustion space = 53 ft^3
g. Projected heating surface = 80 ft^2
h. Average direct contact surface = 30 ft^2
i. Boiler capacity = 5000 lb/hr excluding convection heat transfer; 7000 lb/hr including convection heat transfer
j. 8.75 ft^2 of bed area
k. Heat release rate: 600,000 to 1,200,000 Btu/ft^2hr
l. Pressure rating: 300 psi design, 200 psi normal operating.

4. Air Distribution Grid

The grid contains 815 stainless steel air distribution button spaced on 1¼" centers each containing eight drilled ports, .089" diameter. The air is discharged downward at an angle of 15° to the horizontal.

5. The flue system is fitted first with a water-cooled tube array for temperature quenching, and a two-pass, 104 tube (1" x 6'), 600° F air preheater; this is followed by a second water-cooled gas cooler dust collector, which exists to a 16" duct. The system is drawn by a 4000 cfm, 5" w.g. static pressure, induced draft fan.

6. Dust Collector and Fly Ash Reinjection

   The dust collector contains twelve 10-inch diameter centrifugal collector units with a dust hopper, a 4" Allen-Sherman-Hoff rotary feeder for fly-ash reinjection and a valve for fly-ash removal.

7. Coal Input

   500-900 lbs per hour

8. Thermocouple locations are listed for both the FBM and CBC in Appendix C.

APPENDIX C

CBC SPECIFICATIONS

APPENDIX C. CBC SPECIFICATIONS

1. Air Supply

Shares air supply with FBM from centrifugal fan rated at 2500 cfm, 50" w.g. via air preheater.

Air is supplied via a 4" nominal, Schedule 40 pipe, and is controlled by a gate valve. Flow is orifice monitored.

2. Plenum

Mild steel, ¼" thickness. Approximate dimensions are 10" wide, 10" deep, and 18" high. The plenum rests between the FBM's cross-headers.

3. Column (initial configuration)

A. ASTM 446, stainless steel, ¼" thickness, 10-5/8" x 15-5/8" inside cross-section 56" high.

B. Gas outlet -8" ID diameter starting at 48" level.

C. Front wall contains 18" x 24" access plate.

D. Left wall (facing toward common wall with FBM) contains manometer connections.

E. Right wall (facing toward common wall with FBM) contains thermocouple ports.

F. Back, or common wall contained intercommunication slot(s) with the FBM. Open area, achieved by cutting holes in a steel baffle, varied from 4 square inches to 2 square inches.

4. Air Distribution Grid

Grid contains 96 ASTM 303 stainless steel air distribution buttons spaced on 1¼" centers, each containing eight drilled ports, 0.089" diameter. The air is discharged downward at an angle of 15° to the horizontal.

5. Flue System

From the CBC exhaust, the gas is carried in an 8" pipe to a 4" dust collector unit with a dust hopper, rotary feeder, and a valve for recirculating fly ash, or discharging it to waste.

LOCATION OF THERMOCOUPLES IN FBM/CBC SYSTEM

(Tall CBC Configuration)

NO.	LOCATION*
1.	Inlet air
2.	FBM 1½" bed, 9½"
3.	FBM 1½" bed, 28½"
4.	FBM 1½" bed, 45½"
5.	FBM 1½" bed, 63½"
6.	FBM 9½" bed, 9½"
7.	FBM 9½" bed, 28½"
8.	FBM 9½" bed, 45½"
9.	FBM 9½" bed, 63½"
10.	FBM 21½" bed, 9½"
11.	FBM 21½" bed, 28½"
12.	FBM 21½" bed, 45½"
13.	FBM 21½" bed, 63½"
14.	CBC 1½" bed
15.	CBC 7" bed
16.	CBC 19" bed
17.	CBC 43" bed
18.	CBC 70" overbed
19.	H.V.T. below steam drum (FBM)
20.	H.V.T. above steam drum (FBM)
21.	H.V.T. after convection bank (FBM)
22.	H.V.T. after air preheater (FBM)
23.	H.V.T. after economizer (FBM)
24.	CBC flue gas after tube cooler (superheater
(1-24 recorded on Honeywell)	
25.	CBC: 110" above grid, In Exit Duct, Center, Back
26.	CBC: Below Cooler, In Exit Duct, Center, Back
27.	FBM/CBC: Incoming Cooling Water, All Circuits

28. FBM: Door Cooling, Water and Out
29. FBM: Feedwater to Drum, In
30. FBM: First Gas Cooler, Water Out
31. FBM: Second Gas Cooler, Water Out
32. CBC: Gas Cooler, Water Out
33. FBM: Coal Feeder Cooler, Water Out

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* Location in FBM described as height above grid, distance from front of unit.

APPENDIX D

LABORATORY APPARATUS

APPENDIX D. LABORATORY APPARATUS

The instruments and test apparatus were described in Section 5. Laboratory equipment used in this program are listed below.

1. LECO Resistance Furnace and Sulfur Titrator*, Model 517 w/LECO No. 516 Purifying Train.
2. Coleman Model 33 Carbon-Hydrogen Analyzer.
3. MSA Particle Size Analyzer. (Whitby Centrifuge and projector)
4. Tyler Portable Sieve Shaker and U.S. Standard Sieve Series (Fisher).
5. Thermolyne Muffle Oven
6. Aminco Oven
7. Sartorius 2700 Series Balance
8. Fisher Brand Model 100 Precision Balance
9. OHAUS Triple Beam Balance
10. Boekel Dry Cabinet
11. Beckman Model B Spectrophotometer w/Flame Photometry Attachment and Photovolt Linear/Log Model 43 Recorder.
12. Lindberg tube furnace
13. Norton ball mill.

* Many sulfur analyses of $CaSO_4$ were done by barium titration (Eschka).

APPENDIX E

BACKGROUND INFORMATION ON THE "CHEMICAL ATTRITER PROCESS"

APPENDIX E

BACKGROUND INFORMATION ON THE "CHEMICAL ATTRITER PROCESS"

Among the numerous anomalies which persist in the fluidized-bed combustion of coal is the unexplained difference between limestone effectiveness as the coal is changed, i.e., with a given limestone, Ca/S ratio, reactor, and operating conditions, a change from one coal to another causes a change in SO_2 removal efficiency. The difference cannot be explained on the basis of differences in the total quantity or forms of sulfur. Possibly the most effective demonstration of this anomaly were the results at Coal Research Establishment (CRE) described in Reference E1. Figure 1 of Reference E1 is reproduced here as Figure E-1.

Farmington, a U.S. Coal, and Goldthorpe, a U.K. coal, were both fired in CRE's 6" reactor under identical conditions. Yet all of the sulfur in Goldthorpe could be removed at Ca/S<2.5, while only 89% of Farmington's sulfur could be removed, at the same Ca/S ratio.

A number of reasons have been offered for this difference in results, and apparently a systematic study is under way. A paper at Hueston Woods II, Session II, prepared by Davidson and Smale (E2), suggested that the caking properties of the coal might be used as an indicator of performance, i.e., the sulfur from highly caking coals would not be as easily removed as the sulfur from weakly caking coals. One argument offered as to how this could affect sulfur capture was that the caking coals "envelop" the limestone particles and hence prevent their complete sulfation. While Davidson and Smale have been unable to observe agglomerates, PER has. We note when firing in a fluidized bed of limestone particles that samples withdrawn from the bed contain one or two large particles composed of a number of small limestone particles held together by a central core of coke. We would observe these while CRE would not, since our coal is larger (¼" x 0), i.e. the coupling can only be temporary, and with CRE's fine coal (-1/8" or -1/16") the coke core would be consumed so rapidly that no agglomerates would appear in their sample.

An argument that might be offered then is that the coupled stones are "dead-burned" by conductive heat transfer from the hotter coke particles, or regenerated by contact with hot carbon.

However, we believe that a better insight into the nature of the difference between coals was offered by Henschel (E3). In discussing the difference between Farmington and Goldthorpe, he stated,"...One might, therefore, suspect that some component of the British coal, not present in the American coal, is responsible. Chlorine is one possible candidate, but the culprits could also be some of the compounds of the ash."

The reported chlorine contents (E1) of Goldthorpe (0.2%) and Farmington (0.08%), appear trivial. However, chlorides are highly reactive.

Our argument for the effect of chlorine on lime efficiency is as follows:

> Ash constituents released during fluidized-bed combustion of coal are captured by lime, either physically or chemically (or, first physically and then chemically). One obvious substance is iron; limestone particles in PER's tests with limestone beds show a progressive darkening with time. The color persists after regeneration and is therefore unrelated to the level of sulfate. One likely reaction not only binds up the calcium but releases the sulfur as well.*
>
> $$CaSO_4 + Fe_2O_3 + CO \rightarrow CaO \cdot Fe_2O_3 + SO_2 + CO_2 \qquad (1)$$
>
> The list of potential substances which could react with calcium oxide or sulfate is long, including most every ash constituent, especially the acidic ceramic-forming oxides SiO_2 and M_2O_3. So, rather than a sulfate shell limiting the activity of lime, it would be a shell of

---

*The inference from reactions of this sort is that SO_2 absorption results on lime in gas-fired reactors or in "synthetic flue gas" mixtures do not relate to SO_2 removal capabilities of coal-fired reactors, except as an upper limit to absorption obtainable.

calcium ferrite, silicate, aluminate, etc., i.e, a ceramic shell. Chlorine, in turn, acts to "cleanse" the surface of the particle -- not necessarily by reacting with the ceramic shell, but possibly by undermining it.

To test this hypothesis we analyzed the data of Reference E1 For each run reported we computed the ratio:

$$\text{Ratio} = \frac{\text{Calcium collected in secondary cyclone plus dust}}{\text{Calcium removed in ash or retained in bed}}$$

and plotted the results (See Figure E-2). Some results had to be discarded. (In run 98, for example, the ratio was infinite.) Other problems with the data were poor calcium balances, a leak in the primary cyclones during some of the tests, a large Ca/S ratio (∿0.4) naturally in the Farmington coal. Despite these defects, the results show a trend toward more calcium fines with Goldthorpe (0.2% Cl) than with Farmington (0.08% Cl).

The question which arises is how chlorine in coal leaves the coal and how it attacks the lime particles. A literature search revealed that the nature of chlorine in coal has not been firmly determined, since the methods used to determine its form may themselves change its form. Similarly, the route by which the chlorine is volatilized during combustion is not understood, although it is generally agreed that its final form is HCl.

One path would take it out of the coal as NaCl vapor which would form Na_2SO_4 and HCl via

$$2\ NaCl + H_2O + SO_2 + \tfrac{1}{2}O_2 \rightarrow Na_2SO_4 + 2\ HCl \qquad (2)$$

the basis for the commercial Hargreaves process.

Bench scale experiments at PER have shown that NaCl is capable of dissolving $CaSO_4$ under the conditions existing in a fluidized-bed combustor. The melt has sufficiently low viscosity and surface tension that it does not remain in open containers. The formation of a double salt of calcium and sodium (or potassium) whose crystal properties are sufficiently different from $CaSO_4$, would cause it to "dust" from the lime particle under turbulent fluidized conditions, removing the ceramic shell as well as some of the sulfate shell.

Another route for the action of chlorine in coal on lime would be to form calcium chloride via

$$2\ HCl + CaO \rightleftarrows CaCl_2 + H_2O \qquad (3)$$

in a region of the bed where gas composition and temperature are favorable, and then revert to HCl in another region. Calcium chloride* is molten at the temperature of fluidized bed combustors and can attack silicates.

Since the chlorine is not permanently consumed or converted in reactivating the lime, it acts as if it were a homogeneous catalyst.

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* m.p. 1430°F

It does not appear possible to establish, without ambiguity, the route for coal chlorine attack on partially sulfated lime. However, it seems likely that the weakening of the sorbents' surface, so that it is easily attrited in the fluidized bed, is the likely route. A search for further data comparable to that of Reference E1 was unsuccessful.

Data on calcium in fines are not available for the recent runs with a variety of U.S. and U.K. coals (E4).

What is available from Reference E-4 is a plot of SO_2, ppm in gas versus Ca/S ratio in feed for three coals--Welbeck, Park Mill and Peabody No. 10 (See Figure 5 of Reference E4). Using these data, we computed the moles of calcium per 100 pounds of coal required to lower the emission of sulfur in gas from 1000 ppm to 500 ppm. By this technique we "normalized" the data to equal sulfur contents. We then plotted the result versus the mole ratio S/Cl in the coal (See Figure E-3). Unfortunately, precisely the same curve shape could have been drawn using % sulfur as the abscissa, ignoring chlorine content entirely.

However, since Farmington and Goldthorpe coals have about the same sulfur content, the curve shape cannot be explained on % S alone.

The sulfur and chlorine contents used in this analysis are listed in Table E1 below, along with other coals in the program.

TABLE E1

SULFUR AND CHLORINE CONTENTS OF COALS USED IN FLUIDIZED-BED COMBUSTION TESTS

Coal Name	Place Used	Weight, %		Weight Ratio, S/Cl
		Sulfur	Chlorine	
Arkwright	CRE	2.25	0.09	25
Barington	CRE	0.6	0.14	4.3
Farmington	CRE	2.25	0.08	28
Goldthorpe	CRE	2.05	0.20	10.2
Humphrey	CRE	?	?	?
Park Mill	CRE	2.40	0.14	17.1
Peabody No. 10	CRE	4.06	0.12	34
Peabody No. 10	Argonne	~4.0 to 4.5	?	?
Pitts. No. 8 (Georgetown)	PER	4.73	.05	94
Welbeck	CRE	1.25	0.53	2.4

Hopefully, data will soon replace the question marks and calcium in dust data will also become available, confirming our hypothesis on the action of a coal's chlorine on lime.

Data for Humphrey coal, the fourth coal in CRE's series, is not available from Reference E5. However, from the poor results shown in Figure 1 of this reference for Humphrey coal, it should have a high S/Cl ratio.

The methods used in constructing Figure E-3, and the order-of-magnitude range of the S/Cl ratio, indicate that better methods are required to represent the effect of chlorine on limestone. The difference in calcium fines production versus Ca/Cl ratio is a better tool.

Since we postulate a contamination of the lime by ash matter and a decontamination by the action of chlorine (or a chlorine compound), the differences between the effectiveness of a limestone with a variety of coals should not be described by chlorine content alone. The contaminants have not been identified.

A search of much of the literature on the limestone injection process indicates that while ash contamination of lime has been investigated, none has been identified. However, it may take very little to decrease the reactivity of lime for sulfur.

The following combustion experiments are indicated:

1. For a series of coals reasonably close in sulfur content but with a range of chlorine contents, "normalize" the sulfur contents by the addition of SO_2 as a gas. Test

the coals at one Ca/S ratio based on the normalized sulfur content. Compare the results, SO_2 removal versus chlorine content.

2. Firing gas with SO_2 added, determine the effect of each ash constituent on a fluidized bed of lime. The constituents to be added would be in the mineral form in which they exist in coal, i.e., alumina plus silica as a kaolin, iron as pyrite, etc.
3. Having identified the mineral(s) responsible for lime deactivation in experiment 2, repeat the experiment with a variety of limestones. This is intended to establish if, in fluidized-bed combustion of coal, the difference in apparent reactivity for sulfur is, in fact, the result in a difference in activity for forming a ceramic shell.

While experiments of this nature appear justified if predictive design tools are to become available, it also appeared that a method might be available for reducing the limestone requirement by artificially increasing the chlorine content of the coal.

An important question to be answered is: Will chloride added to the bed as sodium chloride increase the quantity of HCl emitted? Where the sulfur oxide removal process, enhanced by chloride addition, approaches 100%, the Hargreaves process, equation (2), would not be operative. That this may be the case is indicated by the ability of the bed to retain chlorine from the coal when essentially all the sulfur has been captured (E1).

Another question to be answered is the quantity of sodium chloride required to enhance sulfur capture. At 1550°F and high excess air rates (∿700 lb/hr ft^2) about 1 lb/hr ft^2 of NaCl is carried out of the bed as a vapor. At temperatures closer to the melting point of NaCl (1470°F) the NaCl vaporized would be 0.4 lb/hr ft^2 and may be lower in the presence of lime and coal ash (E6).

REFERENCES FOR APPENDIX E:

E1. National Coal Board (CRE); "Retention of Sulphur by limestone in the 0.15 M Fluidised Combustion." February 1969.

E2. Davidson, D.C. and Smale, A.W.; National Coal Board, England; "The Retention of Sulfur by Limestone in a Pilot Scale Fluid-Bed Combustor." Paper II-1 at 2nd International Conference on Fluidized-Bed Combustion, Hueston Woods, Ohio, October 1970.

E3. Henschel, D.B.; Division of Process Control Engineering, NAPCA (now APCO); "Evaluation of British Results Injecting Limestone into a Fluidized-Bed Combustor." February 2, 1970.

E4. National Coal Board, First Three Monthly Report: June 1, 1970 to August 31, 1970; "Reduction of Atmospheric Pollution." Submitted: August 1970.

E5. National Coal Board; "Reduction of Atmospheric Pollution." Monthly Progress Letter for Research on Reducing Emission of Sulphur, Nitrogen Oxides and Particulates by Using Fluidised Combustion of Coal, October 1970.

E6. Wikert, K.; ENERGIE, pp. 12, 240.

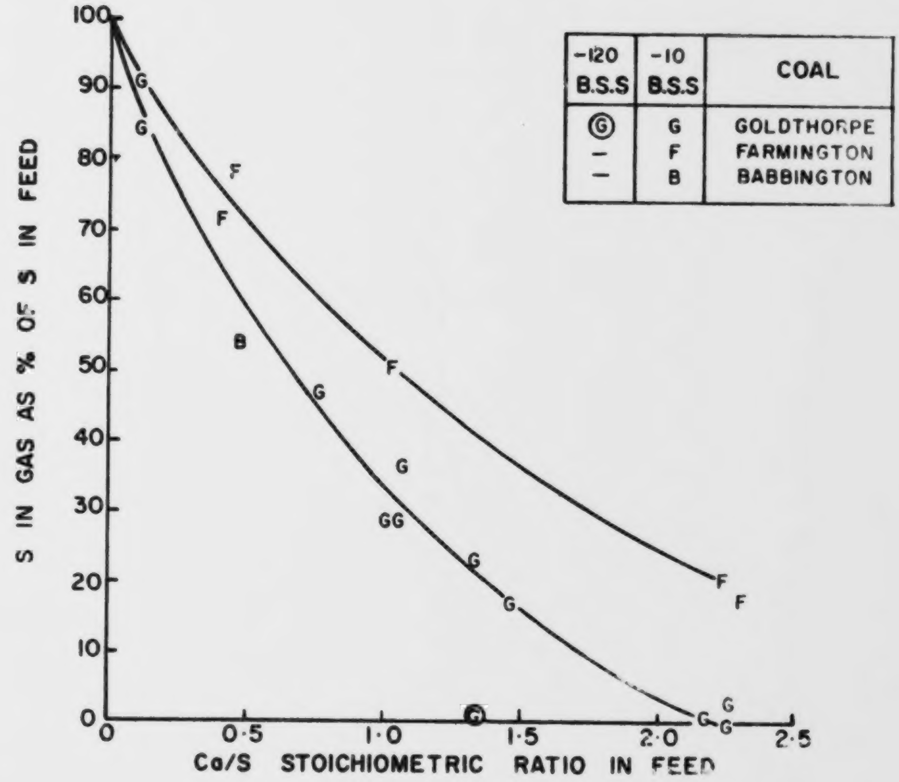


FIGURE E-1. SULFUR RETENTION BY LIMESTONE

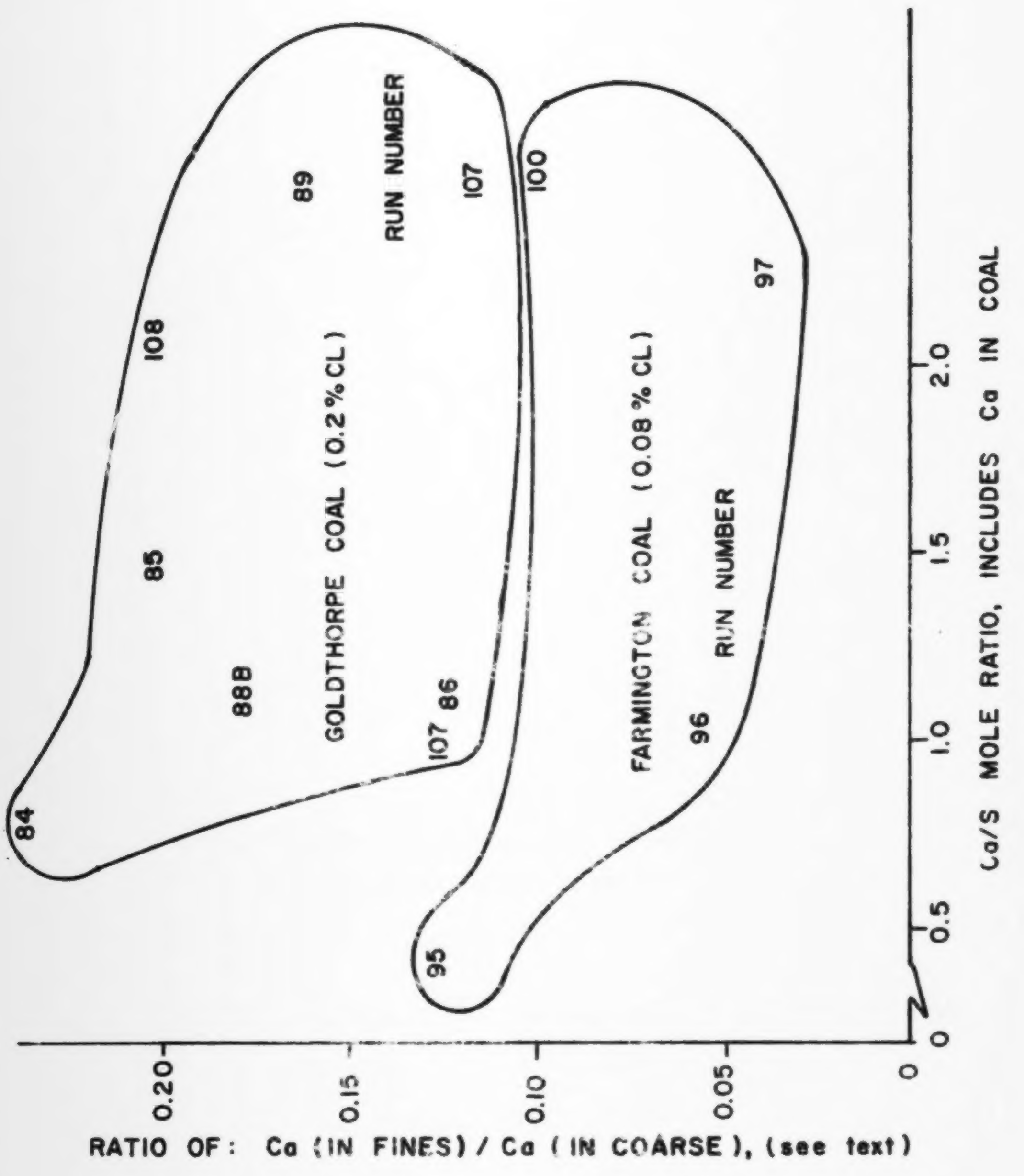


FIGURE E-2. EFFECT OF CHLORINE CONTENT IN COAL ON CALCIUM FINES PRODUCTION IN FLUIDIZED-BED COMBUSTION

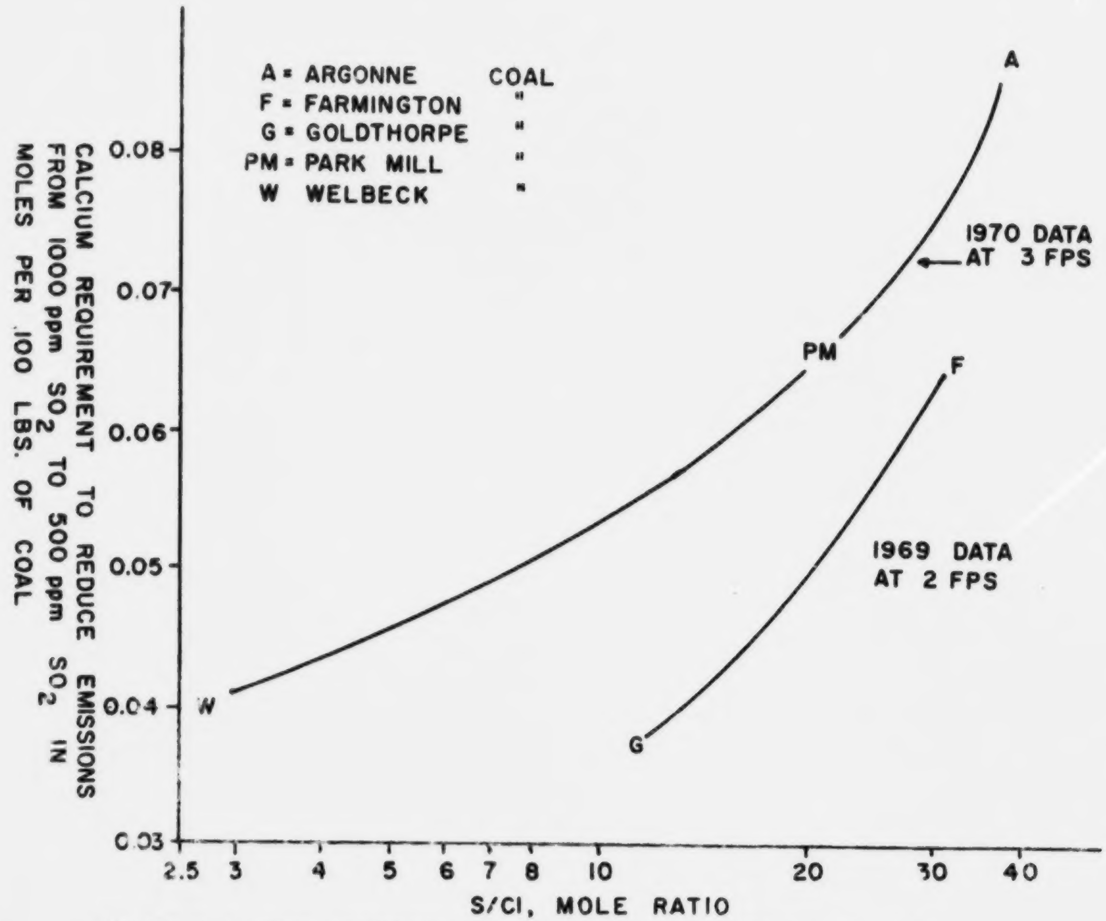


FIGURE E-3. THE EFFECT OF CHLORINE CONTENT OF COAL ON LIMESTONE EFFECTIVENESS IN FLUIDIZED-BED COMBUSTION

APPENDIX F

SULFUR BALANCES

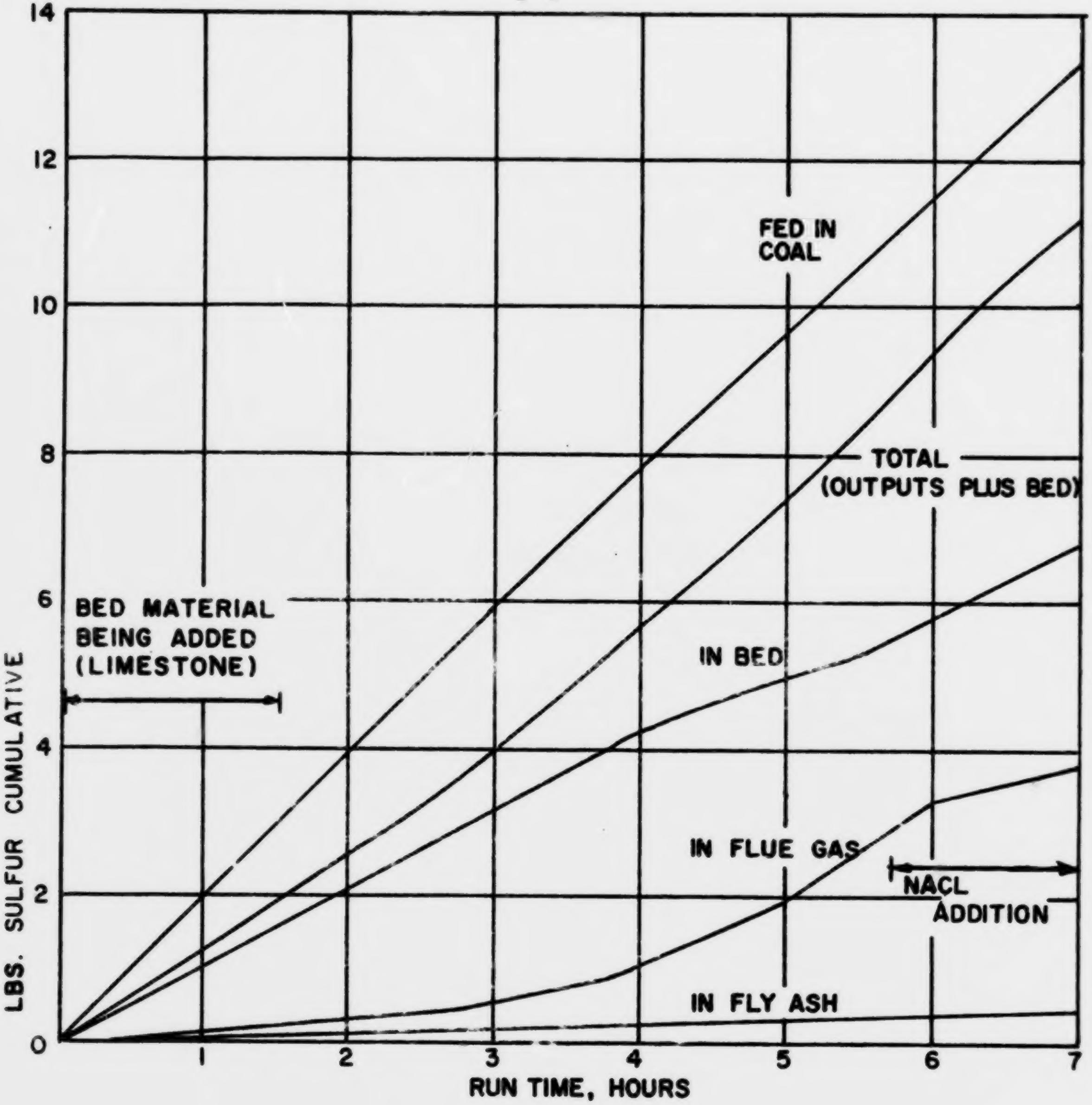


FIG. F-1. CUMULATIVE SULFUR BALANCE, FBC RUN C-321

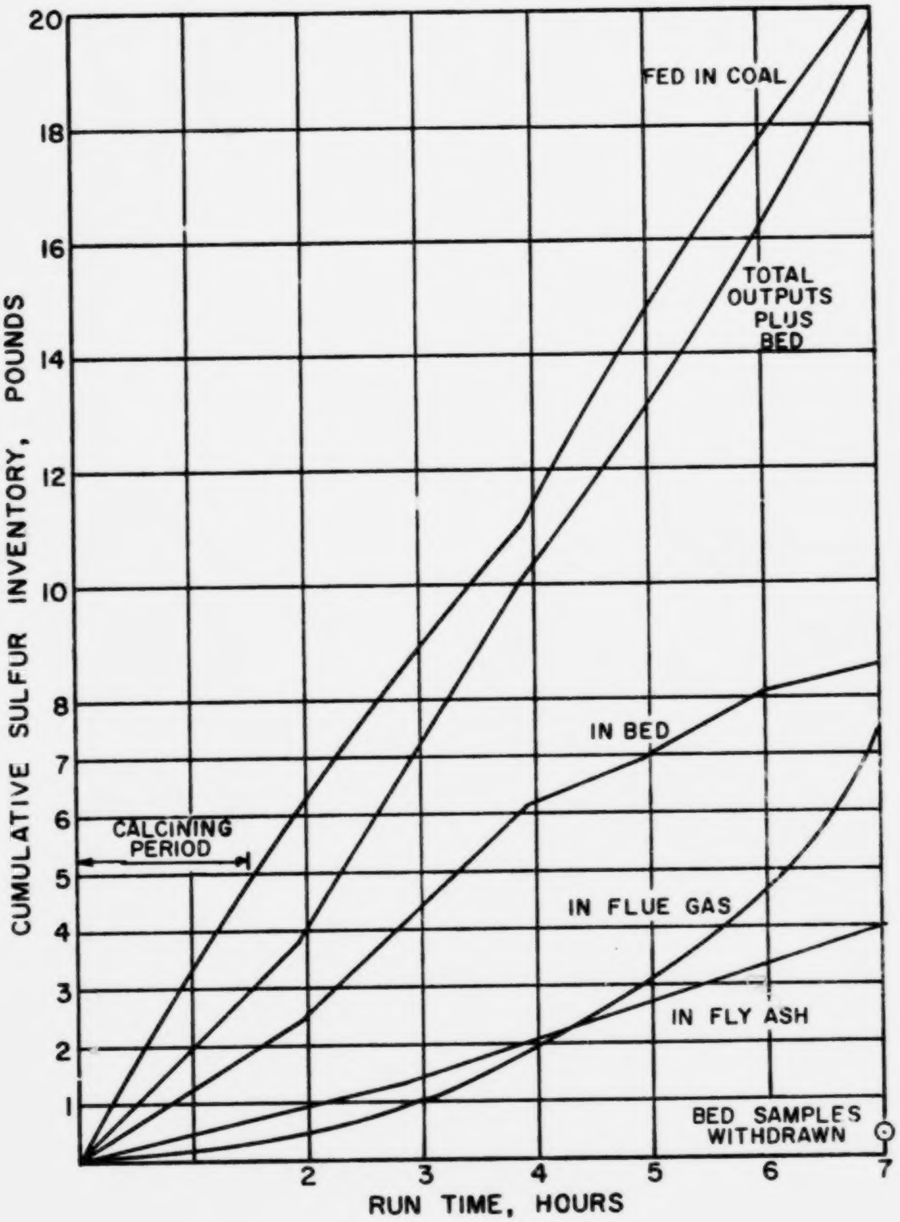


FIGURE F-2. SULFUR BALANCE FOR RUN C-322

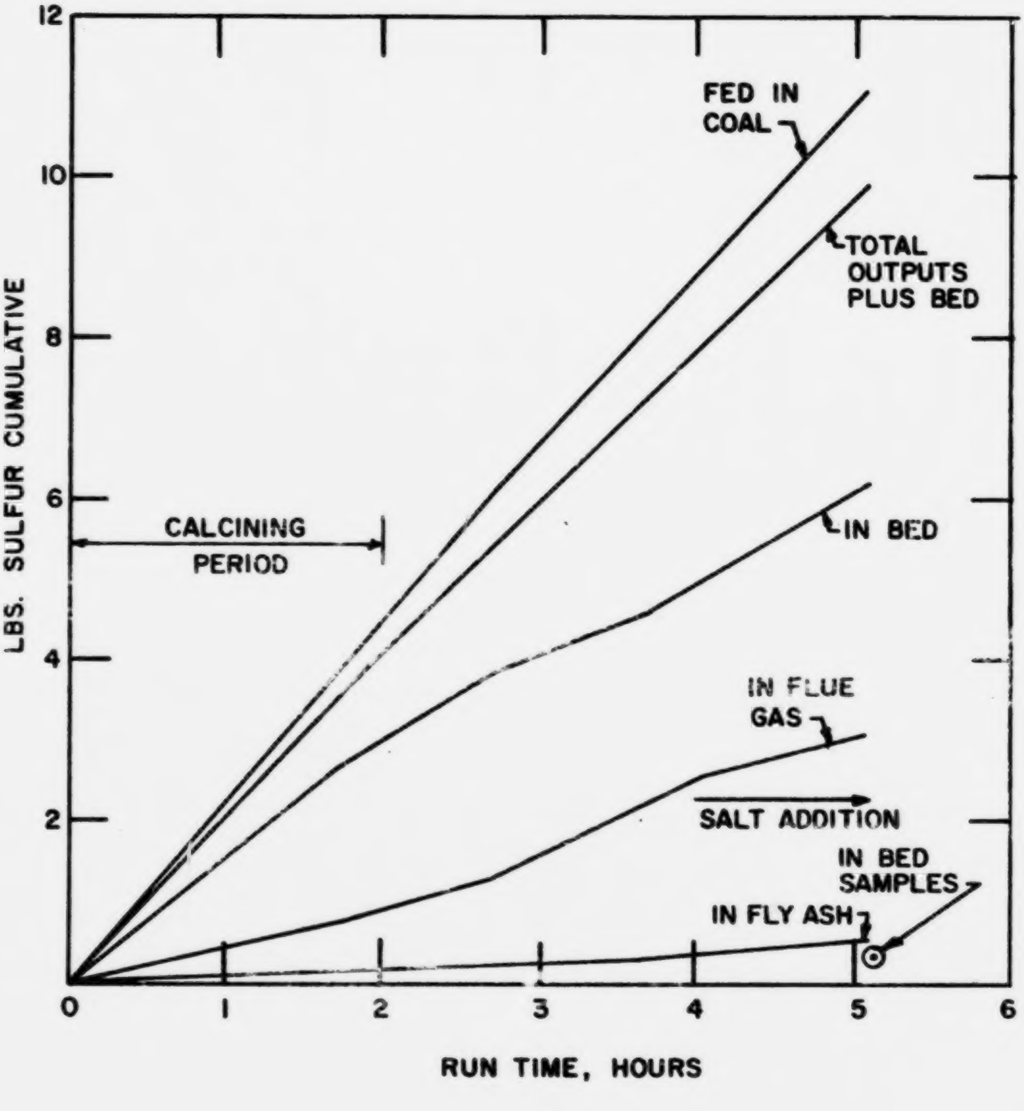


FIGURE F-3. SULFUR BALANCE FBC RUN C-323

TABLE F-1 SULFUR INVENTORY RUN C-324 (Pounds)

Condition	Time	Hours (total)	Fed in Coal, Cum.	In Flue Gas	Cum.	In Fly Ash, Cum.	In Bed Material	% Reduction	Total Accounted For	% Balance
0	8:24	0	0	0	N/A	0	0	N/A	N/A	N/A
1	10:30	2.1	4.35	0.42	0.42	0.40	3.18	90.4	4.00	92
2	11:30	3.1	6.42	0.25	0.67	0.62	4.55	89.6	5.84	91
3	12:30	4.1	8.67	0.38	1.05	0.84	6.05	87.9	7.94	92
(Salt on)	12:50	4.43	N/A	0.17	1.22	N/A	N/A	N/A	N/A	N/A
4	1:30	5.1	11.01	0.28	1.50	1.03	7.27	86.4	9.80	89
5	2:30	6.0*	13.08	0.27	1.77	1.50	9.25	86.5	12.52	96
6	3:30	6.96*	15.01	0.24	2.01	2.15	7.29	86.6	11.45	76
7	4:20	7.8	16.90	0.28	2.29	2.40	10.70	86.4	15.39	91
8	4:38	8.1	17.44	0.17	2.46	2.60	8.19	85.9	13.25	76

*Down time between conditions 4 & 5, 5 & 6 due to salt feed problems, total 8.5 min.

TABLE F-2

SULFUR BALANCE, RUN 168H

				Pounds of Sulfur
A.	Sulfur input to FBM in 4.32% S coal:			
1 a.	March 8	(6 hrs.)		129
1 b.	March 9	(23 hrs.)		546
1 c.	March 10	(24 hrs.)		575
1 d.	March 11	(18 hrs.)		455
			Sub-total	1705
2.	Not captured in FBM:			-130
3.	Input to CBC in coal:			+ 31
			Sub-total	1606
B.	Sulfur in gas output of CBC at avg. 2.5% SO_2			
1 a.	March 8	(6 hrs.)		117
1 b.	March 9	(23 hrs.)		553
1 c.	March 10	(24 hrs.)		381
1 d.	March 11	(18 hrs.)		173
			Sub-total	1224
2.	In bed samples (10 lb/hr x 71 hr x 0.0346)			25
3.	In CBC fly ash (76 lb/hr x 71 hr x 0.0153)			83
4.	In uncollected FBM ash (11.5 lb/hr x 71 hr. x 0.0176)			15
			Total	1347

C. Sulfur unaccounted for: 1606-1347= 259

D. Fraction of total input unaccounted for: 259/1736=0.149

E. Possible error in SO_2 content of CBC Flue gas:
259/(259+1224)=17.5%
(i.e., 2.5% measured - 2.95% actual would explain the imbalance)

TABLE F-3

SULFUR BALANCE, RUN 171-H

		Pounds of Sulfur
A.	Input to FBM in coal:	2440
B.	Input to REG in coal:	180
	Total Inputs:	2620
C.	Emitted as SO_2 from FBM:	270
D.	Emitted as SO_2 from CBC @ 0.27 lb/hr:	43
E.	Emitted as SO_2 from REG (to process):	2107
F.	Lost in CBC fly ash @ 1.54 lb/hr:	240
G.	Lost in FBM uncollected fly ash @ 0.16 lb/hr:	25
H.	In bed material samples (to waste or to scrubber):	55
I.	In bed at end of test:	32
	Total Outputs:	2772
J.	Imbalance: 2772-2620= 152	
K.	% error in REG SO_2 concentration that would account for imbalance: (152/2107 x 100=7%)	

APPENDIX G

PARTICLE SIZE DETERMINATIONS

TABLE G-1

SIZE DISTRIBUTION OF LIME BED (1359), RUN C-321

	Sieve Analysis, Wt. %									
	Time 1.7		3.7		5.5		6.5		7.0 (hr.)	
U.S. Sieve No.	%	Cum.	%	Cum.	%	Cum.	%	Cum.	%	Cum.
+10	19.8		20.6		18.1		18.6		17.2	
-10+14	31.2	80.2	30.0	79.4	32.5	81.9	32.2	81.4	31.9	82.8
-14+18	26.6	49.0	24.9	49.5	26.2	49.4	26.8	49.2	27.6	50.9
-18+20	8.5	22.4	7.5	24.6	7.9	23.2	8.0	22.4	8.8	23.3
-20+25	6.0	13.9	5.4	17.1	5.7	15.3	5.8	14.4	6.1	14.5
-25	7.9	7.9	11.7	11.7	9.6	9.6	8.6	8.6	8.4	8.4
$\overline{d}$, mm*	1.466		1.464		1.451		1.458		1.441	

Apparent attrition rate: 2.2 μm/hr. (radius)

$*\overline{d} = \Sigma \overline{d}_i x_i$

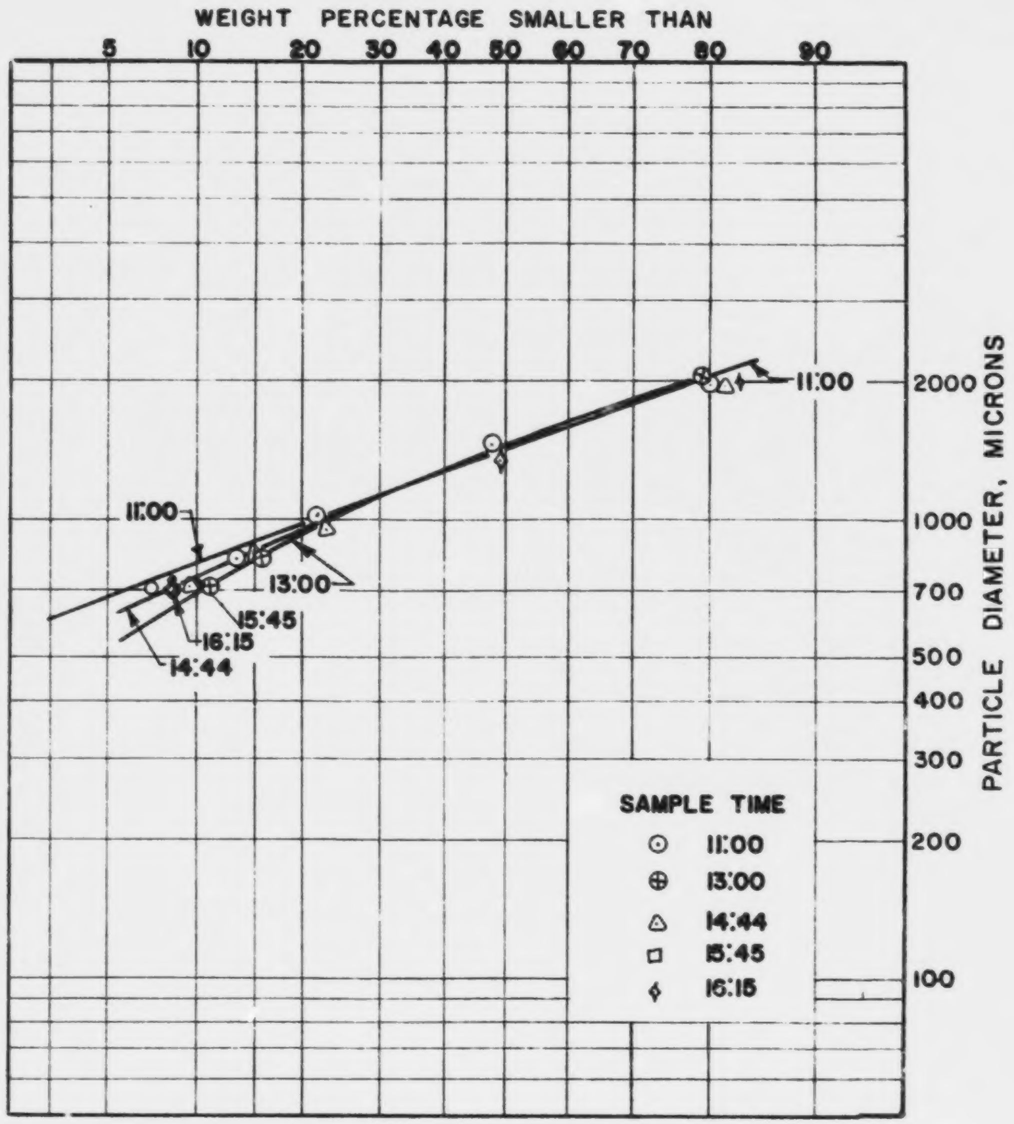


FIGURE G-1. BED PARTICLE SIZE DISTRIBUTION FBC RUN C-321

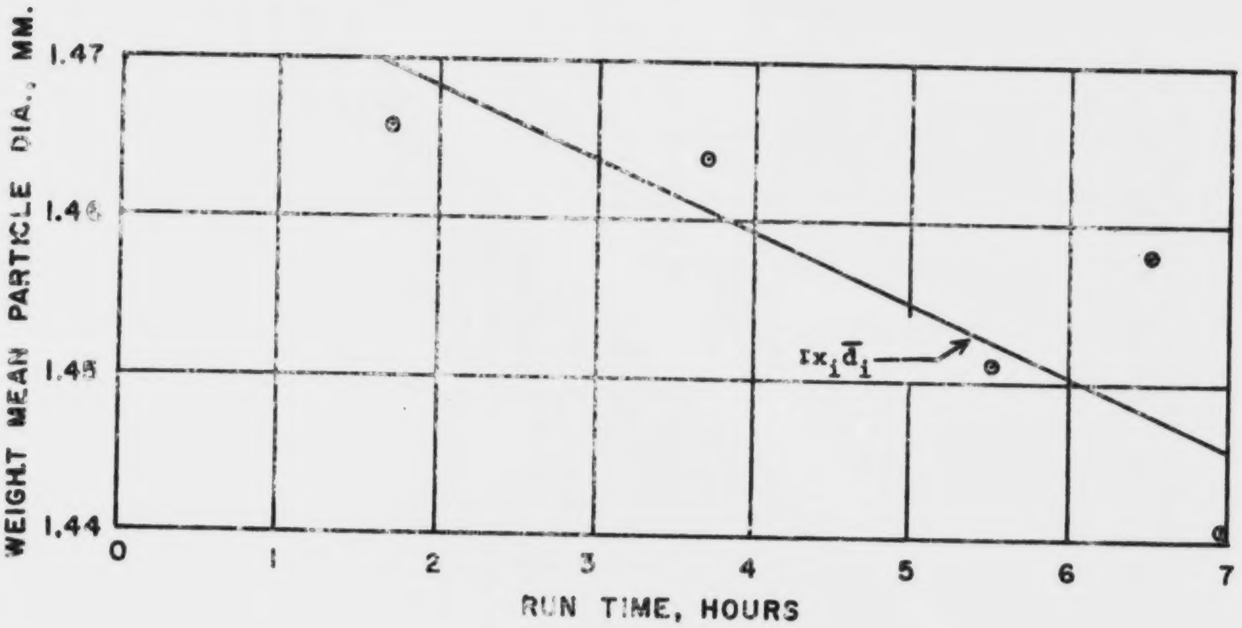


FIGURE G-2. PARTICLE SIZE VS. TIME, RUN C-321

TABLE G-2

SIZE DISTRIBUTION OF LIME BED (1359), RUN C-322

U.S. Sieve No.	Time 2 %	Time 2 Cum.	Time 3 %	Time 3 Cum.	Time 4 %	Time 4 Cum.	Time 5 %	Time 5 Cum.	Time 6 %	Time 6 Cum.	Time 7 (hr) %	Time 7 (hr) Cum.
	Sieve Analysis, Wt. %											
+10	32.1		27.7		29.		25.25		24.6		22.55	
-10+14	24.9	67.84	27.45	72.4	29.2	70.92	27.7	74.87	27.8	75.24	27.7	76.55
-14+18	19.1	42.94	21.8	44.95	21.5	41.72	22.17	47.17	22.8	47.44	22.6	48.85
-18+20	7.17	23.84	8.39	23.15	7.6	20.22	8.0	25.0	8.38	24.64	8.75	26.25
-20+25	5.17	16.67	6.16	14.76	5.15	13.06	6.4	17.0	6.06	16.26	6.5	17.5
-25	11.5	11.5	8.6	8.6	7.91	7.91	10.6	10.6	10.2	10.2	11.0	11.0
Sample wt.,g	541		336		378		385		550		662	
$\overline{d}$, mm	*1.545		1.523		1.553		1.490		1.483		1.446	

Apparent attrition rate: 9 μm/hr (radius)

$*\overline{d} = \Sigma \overline{d}_i x_i$

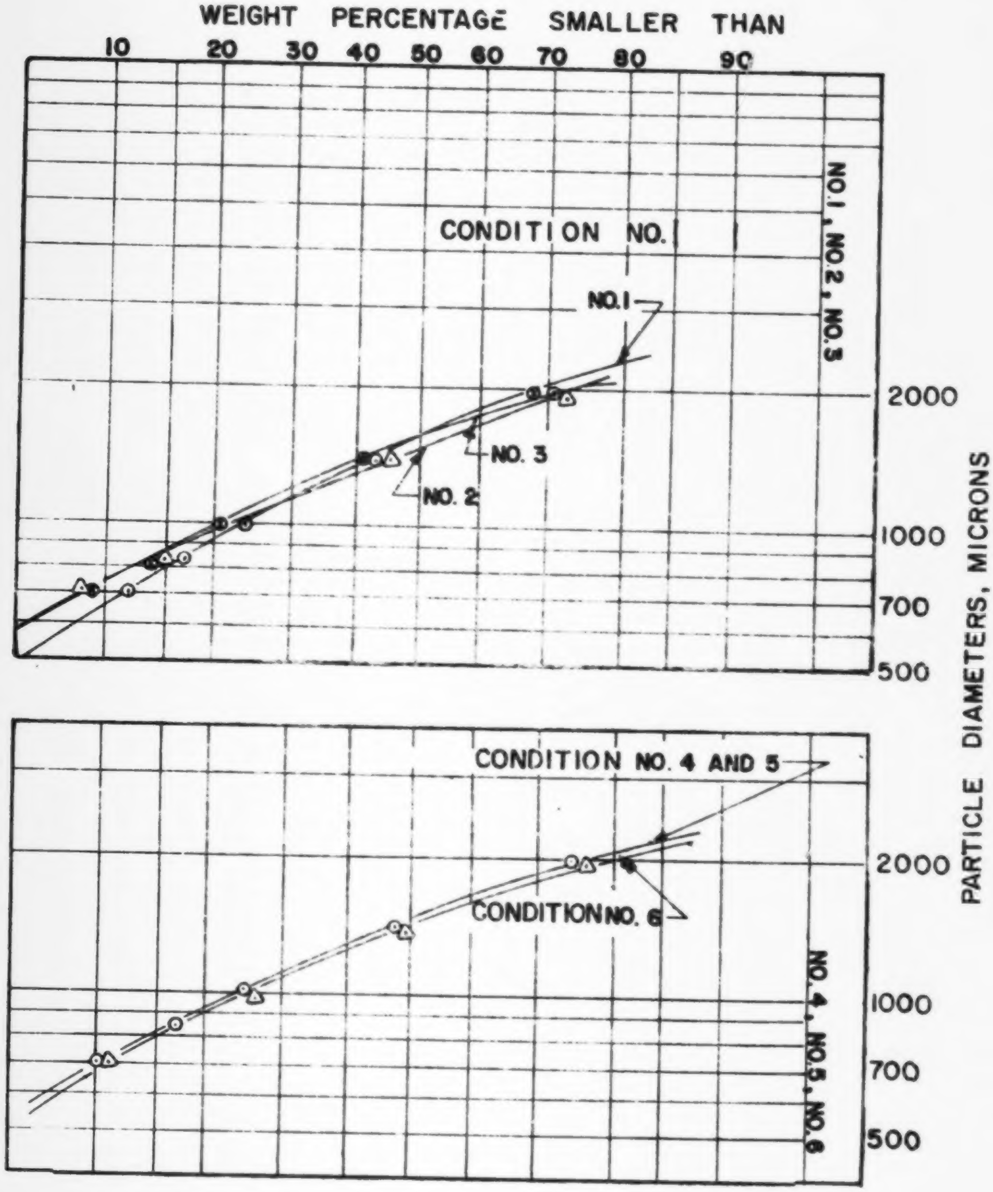


FIGURE G-3, BED PARTICLE SIZE DISTRIBUTION
FBM RUN C-322

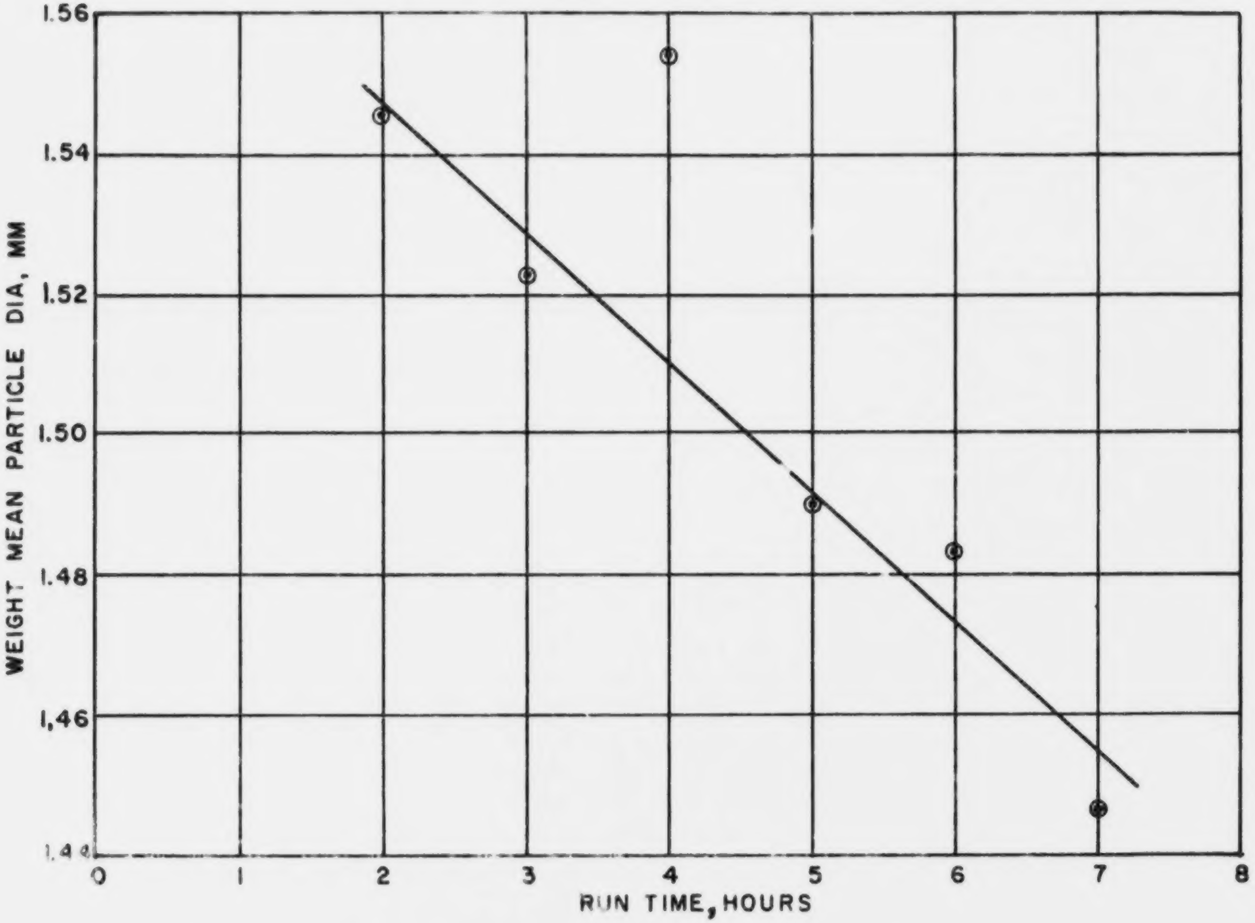


FIGURE G-4. PARTICLE SIZE VS. TIME, RUN C-322

TABLE G-3 SIZE DISTRIBUTION OF LIME BED (1359), RUN C-323
SIEVE ANALYSIS, Wt. %

	Time	1.7		2.7		3.7		4.7		5.1 (hr)	
U.S. Sieve No.	Median, µ	%	cum.	%	cum.	%	cum.	%	cum.	%	cum.
+8	2380	3.42		0		0		0		0	
-8 +10	2190	10.50	96.61	17.40		16.80		17.3		18.75	
-10 +12	1840	10.50	86.11	11.80	82.58	11.78	83.2	11.65	82.7	12.90	81.25
-12 +14	1550	10.40	75.61	10.90	70.73	10.80	71.42	11.32	71.01	12.73	68.35
-14 +16	1300			11.41	59.88	11.51	60.62	11.50	59.69	12.72	55.62
-14 +18	1205	23.30	65.21								
-16 +18	1090			11.82	48.47	12.10	49.11	12.29	48.19	12.35	42.9
-18 +20	920	11.51	41.91	10.20	36.65	10.20	37.01	10.46	35.90	9.61	30.55
-20 +25	774	9.30	30.4	8.55	26.45	8.56	26.81	8.44	25.44	7.34	20.94
-25	700	21.10	21.10	17.90	17.90	18.25	18.25	17.00	17.00	13.60	13.6
$\bar{d}$, mm		1.272		1.329		1.322		1.332		1.385*	
Sample wt., g		505		468		618		778		821	

*Aged sample, some hydration and carbonation are probable.

TABLE G-4 SIZE DISTRIBUTION OF LIME BED (1359), RUN C-324

SIEVE ANALYSIS, Wt. %

Time	2.1		3.1		4.1		5.1	
U.S. Sieve No.	%	cum.	%	cum.	%	cum.	%	cum.
-8 +10	13.41		14.5		14.65		13.95	
-10 +12	11.10	86.61	12.24	85.52	13.28	85.37	12.90	86.03
-12 +14	11.69	75.51	12.45	73.28	13.62	72.09	13.17	73.13
-14 +16	12.00	63.82	12.31	60.83	12.80	58.47	12.64	59.96
-16 +18	12.74	51.82	12.24	48.52	12.84	45.67	12.98	47.32
-18 +20	11.14	39.08	9.82	36.28	10.02	32.83	10.05	34.34
-20 +25	8.29	27.94	7.36	26.46	7.08	22.81	7.42	24.29
-25	19.65	19.65	19.10	19.10	15.73	15.73	16.87	16.87
Sample wt., g	550.6		490.4		560.7		477.8	
$\bar{d}$, mm	1.279		1.308		1.336		1.322	
Sulfur, g	18.7		23.8		36.2		37.1	

TABLE G-4 (Continued)

Time	6.0		6.96		7.8		8.1 (hr)		
U.S. Sieve No.	%	cum.	%	cum.	%	cum.	%	cum.	$\bar{d}$, µm
-8 +10	13.55		11.4		12.3		11.8		2190
-10 +12	12.65	86.43	11.65	88.6	12.95	87.79	14.32	88.2	1840
-12 +14	14.11	73.78	11.95	76.94	13.37	74.84	14.21	73.88	1550
-14 +16	14.11	59.67	12.20	64.99	13.21	61.47	14.05	59.67	1300
-16 +18	14.00	45.56	12.63	52.79	13.21	48.26	13.50	45.62	1090
-18 +20	9.67	31.56	10.65	40.16	10.15	35.05	10.21	32.12	920
-20 +25	7.29	21.89	8.01	29.51	7.82	24.90	7.16	21.91	774
-25	14.60	14.60	21.50	21.50	17.08	17.08	14.75	14.75	"700"
Sample wt., g	712		527.1		536.4		546		
$\bar{d}$, mm	1.330		1.257		1.304		1.321		
Sulfur, g	70.2		40.9		51.1		44.2		322

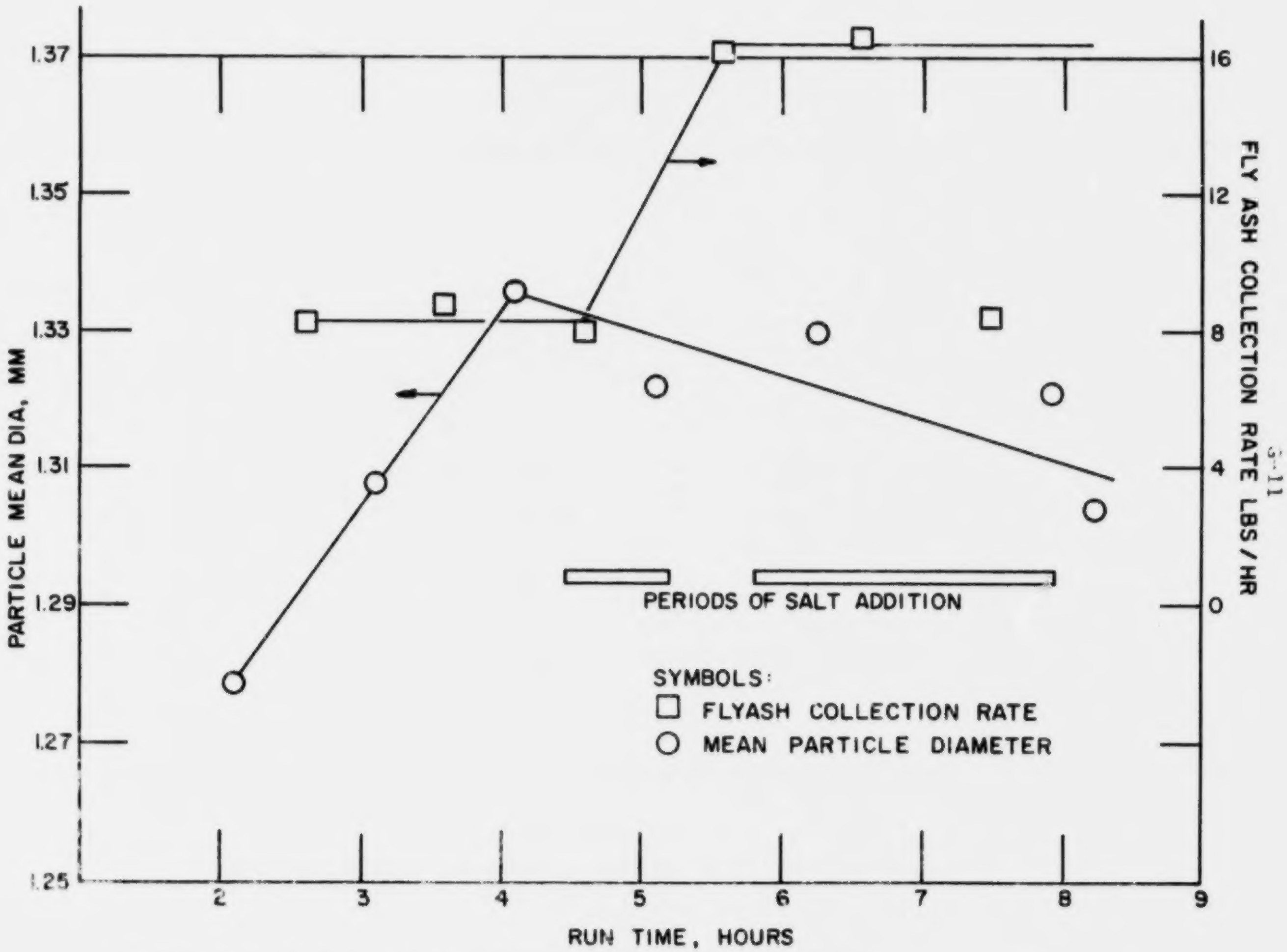


FIGURE G-5 BED PARTICLE SIZE VS TIME, FBC C-324

TABLE G-5 SIZE DISTRIBUTION OF LIME BED (1359), RUN B-18
SIEVE ANALYSIS, Wt. %

Time	3.9		7.7		9.4		10. (hr)	
U.S. Sieve No.	%	cum.	%	cum.	%	cum.	%	cum.
-8 +10	13.2		10.0		7.36		10.5	
-10 +12	11.30	86.9	10.67	90.1	9.21	92.7	11.61	89.6
-12 +14	14.80	75.61	15.70	79.43	15.10	83.47	16.38	78.01
-14 +16	16.48	60.81	18.60	63.73	18.55	68.37	19.23	61.33
-16 +18	20.20	44.33	20.05	45.13	21.24	49.82	19.28	42.40
-18 +20	12.91	24.13	13.75	25.08	14.34	28.58	12.27	23.12
-20 +25	6.30	11.22	5.93	11.33	6.84	14.24	5.47	10.85
-25	4.92	4.92	5.40	5.40	7.40	7.40	5.38	5.38
$\bar{d}$, mm	1.393		1.329		1.274		1.346*	
Sample wt., g	551		287		304		544	
Sulfur content, wt., %	3.12		5.95		6.82		5.98	
Calcium content, wt., %	56.7		54.0		54.4		56.9	

*After partial regeneration (See Figure 4).

TABLE G-6 SIZE DISTRIBUTION OF LIME BED (1359). RUN 168H

	Sieve Analysis, Wt.%						
	FBM Day 8 Time 1949			FBM Day 9 Time 1115		FBM Day 9 Time 1335	
U.S. Sieve No.	Wt. %	Cum.	$\bar{d}_i$	%	Cum.	%	Cum.
+10	2.2		2.19	1.2		1.4	
-10+12	1.3	97.8	1.84	1.1	98.7	1.2	98.6
-12+14	2.4	96.4	1.55	2.4	97.6	2.3	97.2
-14+16	6.3	94.	1.3	6.5	95.2	6.1	94.9
-16+18	14.8	87.7	1.09	13.8	88.7	14.3	88.8
-18+20	21.9	72.9	0.92	23.2	74.9	21.1	74.5
-20+25	17.7	51.	0.774 0.718	35.	51.7	18.4	53.4
-25+30	17.7	33.3	0.651			18.4	35.
-30	15.6		0.55	16.7		16.6	
Sample Wt. g.	142.7			130.4		144.3	
$\bar{d}$, mm. *	0.891			0.873		0.871	
% S	7.84			6.88		7.56	
Air Rate lb/ft²hr	680			760		725	
T °R	2000			1940		1945	

* $\bar{d}_i x_i$

TABLE G-6 Continued

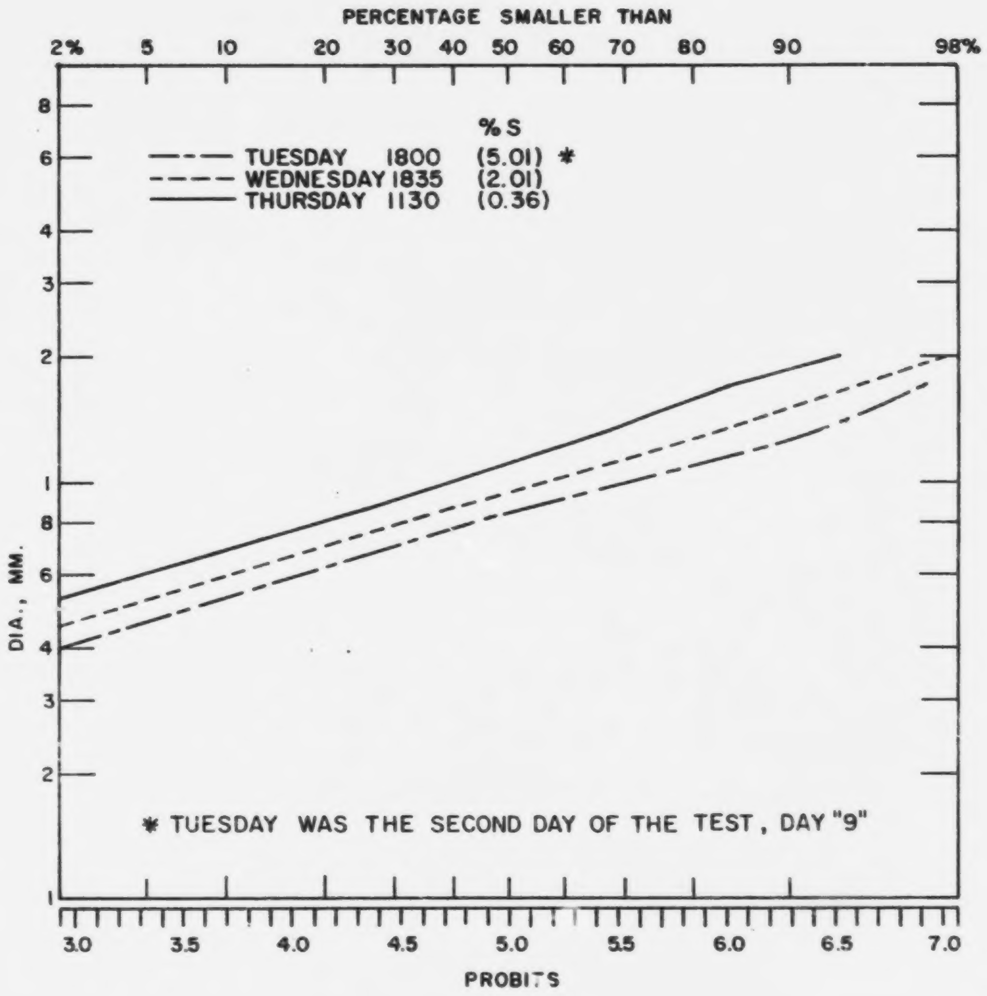
U.S. Sieve No.	CBC Day 9 Time 1800 Wt.%	Cum.	FBM Day 9 Time 1910 Wt %	Cum.	FBM Day 10 Time 0900 Wt %	Cum.
	Sieve Analysis, Wt.%					
+10	1.2		1.9		1.8	
-10+12	1.4	98.8	1.3	98 2	2.4	98.2
-12+14	2.8	97.2	2.7	96.9	4.2	95.8
-14+16	7	94.4	6.6	94.2	10.	91.6
-16+18	15.8	87.4	15.3	87.6	17.9	81.6
-18+20	22.2	71.6	21.9	72.3	22.7	63.7
-20+25	33.1	49.4	19.	50.4	17.8	41.
-25+30			17.6	31.4	14.1	23.2
-30	16.3		13.8		9.1	
Sample Wt. g.	129.7		127.2		115.6	
$\bar{d}$, mm.	0.890		0.900		0.962	
% S	5.01		3.96		1.26	
Air Rate lb/ft^2hr	815		815		790	
T °F	2380		2010		1985	

TABLE G-6 Continued

U.S. Sieve No.	Sieve Analysis, Wt.%					
	FBM Day 10 Time 1409		CBC Day 10 Time 1835		FBM Day 10 Time 1945	
	Wt. %	Cum.	Wt. %	Cum.	Wt. %	Cum.
+10	5.5		2.0		4.2	
-10+12	7.4	94.5	3.8	98.	4.0	95.8
-12+14	11.5	87.	6.6	94.2	6.9	91.7
-14+16	15.4	75.5	11.5	87.6	12.	84.8
-16+18	18.2	60.1	18.	76.1	17.4	72.8
-18+20	18.	41.9	21.1	58.1	19.1	55.4
-20+25	11.7	23.9	28 }	37.	27.8 }	36.3
-25+30	7.9	12.2				
-30	4.3		9.		8.5	
Sample WT.g.	125.5		129.		131.3	
$\bar{d}$, mm.	1.164		1.007		1.042	
% S	1.4		2.01		2.37	
Air Rate lb/ft^{2}hr	775		850		785	
T °F	1990		2510		1980	

TABLE G-6 Continued

	Sieve Analysis, Wt.%					
	CBC Day 11 Time 0330		FBM Day 11 Time 0835		CBC Day 11 Time 1130	
U.S. Sieve No.	Wt. %	Cum.	Wt. %	Cum.	Wt. %	Cum.
+10	1.9		2.7		6.	
-10+12	2.1	98.1	1.8	97.1	10.	94.
-12+14	3.6	96.	3.7	95.3	10.9	83.7
-14+16	8.2	92.4	8.6	91.6	14.1	72.6
-16+18	15.1	84.2	16.9	83.	17.7	58.7
-18+20	17.9	69.1	20.3	66.1	16.7	41.
-20+25	16.	51.2	30.9	45.8	20.1	24.3
-25+30	16.4	35.2				
-30	18.8		14.9		4.2	
Sample Wt., g.	124.9		118.3		114.3	
$\bar{d}$, mm.	0.908		0.936		1.181	
% S	0.7		1.18		0.36	
Air Rate lb/ft^{2}hr	580		780		870	
T °F	2360		2000		2435	



* TUESDAY WAS THE SECOND DAY OF THE TEST, DAY "9"

FIGURE G-6, PARTICLE SIZE DISTRIBUTION OF CBC BED-RUN 168H

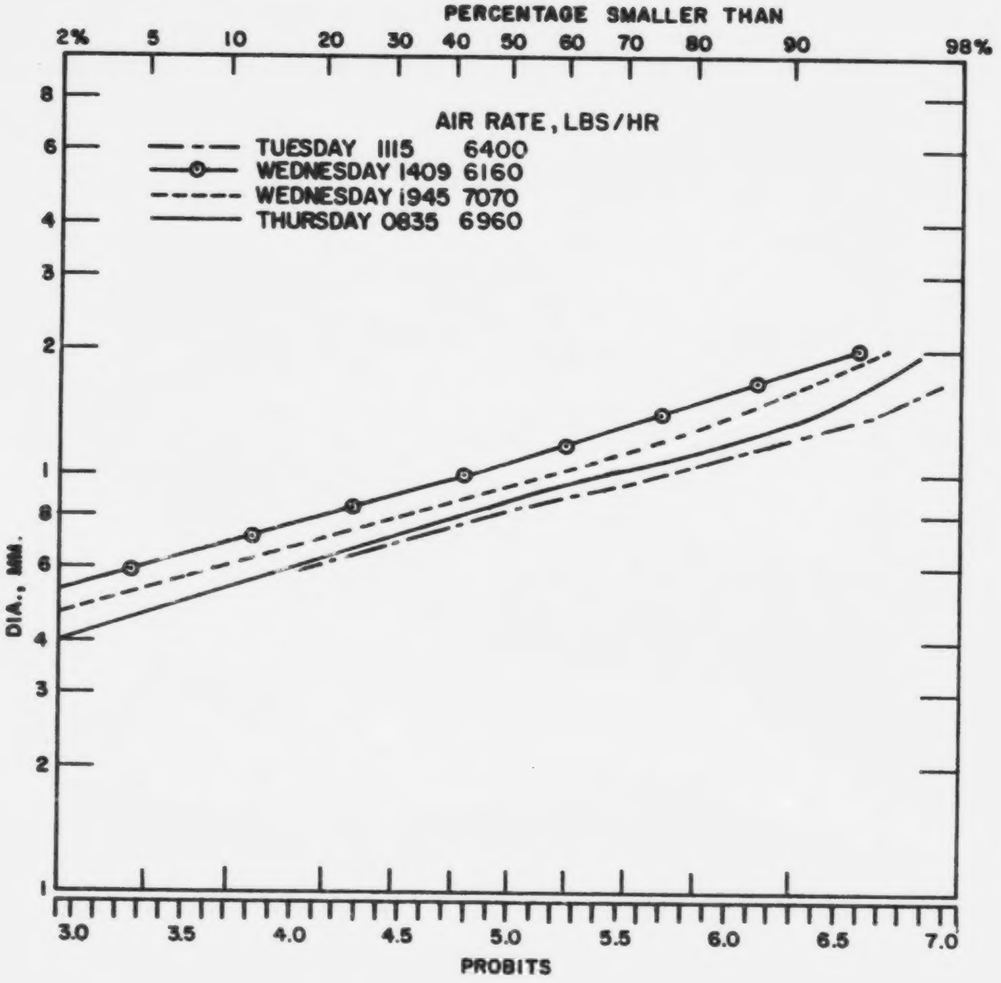


FIGURE G-7. PARTICLE SIZE DISTRIBUTION OF FBM BED - RUN 168H

TABLE G-7 Bed Size Distribution FBM Extended Run July 197

DATE & TIME	7/12	2245	7/13	0845	7/14	2250	7/15	1245	7/16	1000	7/17	0900	7/18	2100	7/19	1100	
Mesh Size*	Retained	Cumulative	Retained	Cumulative	Retained	Cumulative	Retained	Cumulative	Retained	Cumulative	Retained	Cumulative	Retained	Cumulative	Retained	Cumulative	$\bar{d}_i$ Microns
+10	3.82	99.99	3.66	100.00	4.75	99.99	3.80	100.00	3.43	99.99	4.04	100.00	6.15	99.99	5.17	99.99	2190
+12	5.90	96.17	5.58	96.34	5.32	95.24	4.44	96.21	4.21	96.56	3.30	95.36	5.10	93.84	4.89	94.82	1840
+14	7.38	90.27	6.93	90.76	6.46	89.92	5.56	91.77	5.32	92.35	4.12	92.66	5.19	88.74	5.80	89.93	1550
+16	13.95	82.89	13.25	83.83	11.99	83.46	11.49	86.21	10.94	87.03	9.58	88.54	10.69	83.55	14.43	84.13	1300
+18	17.77	68.94	17.31	70.58	16.23	71.47	16.45	74.72	15.45	76.09	17.40	78.96	10.50	72.86	21.86	69.70	1090
+20	20.97	51.17	20.33	53.27	20.31	55.19	20.44	58.27	21.55	60.64	22.64	61.56	22.59	54.36	23.24	47.84	920
+25	14.24	30.20	15.55	32.94	14.47	34.88	15.58	37.83	15.69	39.09	17.18	38.92	16.32	31.77	15.21	24.60	774
+30	N/A	N/A	N/A	N/A	10.54	20.41	13.74	22.25	14.17	23.40	14.91	21.74	11.52	15.45	7.38	9.39	651
+120	15.96	N/A	17.39	N/A	9.87	N/A	8.51	N/A	9.23	N/A	6.83	N/A	3.93	N/A	2.01	N/A	500
Total Sample Weight, grams	421.2		391.0		193.5		433.1		460.7		454.1		325.4		398.4		
$\bar{d}$, mm	1.076		1.059		1.052		1.020		1.003		0.999		1.078		1.108		

* U.S. Sieve Number

APPENDIX H

FBM RUN B-18 LOG

APPENDIX H

TABLE H-1. FBM RUN B-18 LOG

		FLUE GAS COMPOSITION: (FBM unless stated CBC)					
Time	SO_2, ppm	NO	HC	O_2, %	CO_2	T, °F	Coal Rate Lb/hr
8:51	5	N/A	N/A	5.5	N/A	1700	N/A
8:57	300	N/A	N/A	5.5	N/A	1700	N/A
9:03	250	N/A	0	5.4	N/A	1630	N/A
9:09	250	N/A	0	5.5	N/A	1650	610
9:15	300	N/A	0	5.8	N/A	1630	640
9:21	300	N/A	0	6.	N/A	1650	N/A
9:27	300	N/A	0	5.8	N/A	1600	N/A
9:33	325	240	0	5.7	13.5	1590	N/A
9:39	250	250	0	5.5	13.5	1600	N/A
9:45	720	260	0	5.5	13.4	1720	N/A
9:51	610	150	0	3.5	16.	1610	N/A
9:57	930	125	0	3.	16.5	1620	N/A
10:03	490	140	0	2.2	16.1	1570	N/A
10:09	330	140	0	2.9	16.	1550	N/A
10:15	280	110	0	2.7	16.	1520	N/A
10:21	0 (CBC)*	0 (CBC)	0	2.5 (FBM)	0. (CBC)	1480 (FBM)	N/A
10:27	150 (FBM)	80 (FBM)	0	3.	14. (FBM)	1510	N/A
10:33	200	90	0	2.	17.1	1500	N/A
10:39	250	120	0	1.3	17.	1510	N/A
10:45	325	110	0	1.4	17.	1520	N/A
	Increase air						
10:51	200	180	0	3.	15.5	1510	N/A
10:57	225	185	0	3.2	15.4	1520	N/A
11:03	275	180	0	3.	15.	1550	N/A
11:09	320	200	N/A	3.	14.4	1570	N/A
11:15	520	180	900	3.3	14.6	1600	728
11:21	660	170	1200	3.	14.6	1610	N/A
11:27	780	170	N/A	2.8	14.7	1620	N/A

* Notation "(CBC)" means this entry is a CBC flue gas determination.

TABLE H-1. (Continued)

Time	SO₂, ppm	NO	HC	O_2, %	CO_2	T, °F	Coal Rate, Lb/hr
11:33	800	190	180	2.7	14.5	1620	
11:39	700	220	240	2.8	13.8	1620	
11:45	700	220	120	3.5	13.7	1620	
11:51	720	220	60	3.8	13.8	1620	
11:57	750	220	50	3.7	13.8	1615	
12:03	780	220	57	3.7	13.8	1620	
12:09	850	220	120	3.7	14.5	1620	
	Add limestone						
12:15	930	180	120	3.5	14.5	1625	
12:21	1000	220	135	3.	14.	1615	
12:27	460	140	120	3.8	14.	1490	
12:33	460	110	160	3.8	14.4	1470	
12:39	460	100	220	3.4	14.	1470	
12:45	490	100	340	2.8	15.6	1460	
12:51	580	110	460	2.6	15.7	1495	
12:57	640	120	480	2.6	15.7	1505	
1:03	N/A	0 (CBC)	600	2.4	0. (CBC)	1510	
1:09	0 (CBC)	220 (FBM)	0 (CBC)	3.	15.1 (FBM)	1515	
1:15	700 (FBM)	250	90	3.	15.	1520	
1:21	780	270	300 (FBM)	3.	14.9	1540	711
1:27	880	290	360	3.	14.5	1560	
1:33	1000	290	300	3.4	14.3	1575	
1:39	1200	290	300	3.2	12.8	1600	
1:45	940	350	180	3.	12.4	1575	
	Initiate CBC operation						
1:51	1030	320	47	3.6	12.5	1590	
1:57	1260	320	15	3.6	13.6	1605	
2:03	1550	320	150	3.4	13.7	1620	
2:09	N/A	N/A	~120	3.5	N/A	1650	

TABLE H-1. (Continued)

Time	SO_2, ppm	NO	HC	O_2, %	CO_2	T, °F	Coal Rate, Lb/hr
2:15	1800	320	60	3.	13.7 (FBM)	1650	
2:21	1950	320	150	3.4	13.6	1660	
2:27	1900	320	150	3.5	13.	1650	
2:33	1350	330	120	3.5	-	1610	
2:39	1400	305	15	4.8	12.8	1610	
2:45	1400	310	160	4.	12.	1600	503
2:51	1200	320	100	4.5	12.5	1580	
2:57	1050	340	80	5.	11.7	1550	
3:03	1000	330	40	5.5	11.8	1530	480
3:09	6450 (CBC)	470 (CBC)	300 (CBC)	5.5	18.4 (CBC)	1790 (CBC)	
3:15	1200 (FBM)	330 (FBM)	130 (FBM)	6.	12.4 (FBM)	1530	
3:21	1200	330	70	5.5	12.7	1560	
3:27	1230	330	40	4.6 (FBM) 4.5 (CBC)	12.6	1580	
3:33	1200	340	30	5.5 (FBM) 3. (CBC)	12.2	1590	450
3:39	1100	340	16	5.5 (FBM) 3. (CBC)	14.1	1590	
3:45	15,500 (CBC)	240 (CBC)	<16	5.5 (FBM) 2. (CBC)	18. (CBC)	1600	
3:51	-	360 (FBM)	<16	5. (FBM) 10. (CBC)	13. (FBM)	1610	
3:57	1430 (FBM)	370	<16	5. (FBM) 1.3 (CBC)	12.8	1610	
4:03	1260	370	<16	4.8 (FBM) 1.8 (CBC)	12.3	1610	
4:09	1070	340	<16	5. (FBM) 3. (CBC)	12.2	1600	

TABLE H-1. (Continued)

Time	SO_2, ppm	NO	HC	O_2, %	CO_2	T, °F	Coal Rate, Lb/hr
4:15	900	350	<16	5.5 (FBM) 3. (CBC)	12.	1590	
4:21	900	350	<16	5. (FBM) 1.5 (CBC)	12	1600	
4:27	940	380	<16	4.2 (FBM) 0. (CBC)	12.6	1620	
4:33	1040	380	<16	4.	13.2	1620	
4:39	24,400 (CBC)	260 (CBC)	<16	4.	14. (FBM)	1630	
4:45	25,900 (CBC)	380 (FBM)	<16	4.2	17.9 (CBC)	1880 (CBC)	
4:51	1700 (FBM)	400	<16	4.	13.2 (FBM)	1625	
4:57	1300	360	<16	4.5	13.	1650	
5:03	2200	330	<16	2.	15.	1695	
Shutdown							
5:09	2350 (FBM) 7700 (CBC)	330	<16	10.	15.	1700	
5:15	1700 (CBC)	0	0	20.	0. (FBM)	1600	
5:21	1400 (CBC)	0	0	|	0. (CBC)	1400	
5:26	1100 (CBC)	0	0	↓	0.	1300	

Post shutdown SO_2 evolution in CBC (20% O_2) is interesting.

APPENDIX I

RUN 168H - CONDENSED DATA - CO AND HYDRO CARBON EMISSIONS

APPENDIX I

Run 168H

Condensed Data - CO and Hydrocarbon Emissions

1. CBC Carbon Monoxide (in regeneration mode)

   35 measurements, average 0.54%
   median 0.36%

2. CBC hydrocarbon

   60 measurements, average 80 ppm
   median 46 ppm

   These low values are attributed to the high CBC (refractory) freeboard. The intense radiation field chews up HC very effectively.

3. FBM Carbon monoxide - no correlation with Temperature (1450 to 1590°F)

   39 measurements, average 0.53%
   median 0.23%

   CO heating value loss less than carbon loss, on FBM O_2 levels are usually higher then CBC/Regen. O_2 levels. The need for further coal feeder development is apparent.

4. FBM hydrocarbon

   65 measurements, average 630 ppm
   median 400 ppm

   HC heat loss 1/3 of carbon loss. The present coal feeder is not optimized for low hydrocarbon emission.

APPENDIX J

BED CHARACTERISTICS, RUN 168H

BED CHARACTERISTICS, RUN 168H

Day	Time	Reactor	%S	Sp. G.*	d̄,mm	CBC Flue gas Sulfur lb/hr	Sp. G Ratio F/C	S Ratio F/C
8	2015	F	7.84	1.442 ±.009	0.84	8.5		
9	0255	F	6.79	1.421	(graphical)			
		C	6.49	1.259			1.13	1.05
	0430	F	2.50	1.225				
		C	3.18	1.181			1.04	
	0610	F	3.68	1.256				
		C	4.93	1.233			1.02	
	0720	F	4.95	1.297				
		C	5.31	1.240			1.04	
	0840	F	5.60	1.303		25.4		
	0949	F	6.20	1.349				
		C	5.44	1.306			1.03	1.14
	1051	F	6.88	1.310 ±.015	0.82	13.1		
		C	6.62					1.04
	1152	F	8.15	1.375				
		C	6.84	1.366			1.01	1.19
	1345	F	7.56	1.450	0.81	3.7		
		(C)	(7.48)	(1.371)			1.05	1.01
	1655	F	7.80			10.3		
	1800	F	6.38			28.8		
		C	5.01					1.27
	1910	F	3.96	1.313 ±.010	0.84	42.7		
	2225	F	2.6	1.289	0.84	43.7		
		(C)	(2.2)					1.18
	2300	F	2.82	1.172				
		C	2.47	1.292			0.91	1.14
10	0500	F	1.40	1.175		22.6		
		C	0.80	1.236			0.95	1.75
	0635	F	1.41	1.144		6.		
		C	1.48	1.211		10.5	0.95	~1.

---

*Determined by weighing the net contents of a container of known volume poured without tapping.

APPENDIX J
Continued

Day	Time	Reactor		%S		Sp. G		d̄, mm	CBC Flue gas Sulfur lb/hr	Sp. G Ratio F/C	S Ratio F/C
10	0821	F		1.24		1.197				1.05	1.9
			C		0.65		1.143		10.3		
	0900	F		1.26		1.163	±.002	0.91		0.98	2.6
			C		0.48		1.192		9.		
	1029	F		1.16		1.180					
	1128	F		1.07		1.176					
	1200	F		1.11		1.211					
	1342	F		1.19		1.274			12.3		
	1409	F		1.45		1.260	±.006	1.07	12.3		
	1542	F		1.88		1.312					
	1646	F		3.98		1.348				1.05	1.19
			C		3.35		1.278		11.8		
	1758	F		3.56		1.367					
			C		4.58				31.9		
	1857	F		3.03							1.5
			C		2.01		1.307	0.93	29.		
	1953	F		2.37		1.331		0.95			
	2250	F		1.48					16.1		2.5
			C		0.58						
	2345	F		1.47							2.5
			C		0.57		1.229		18.3		
11	0315	F		1.36		1.332				1.05	1.9
			C		0.70		1.264	0.83	9.8		
	0645	F		1.20							1.8
			C		0.66				17.1		
	0830	F		1.18		1.198		0.87			2.2

APPENDIX J
Continued

Day	Time	Reactor	%S		Sp. G		$\bar{d}$, mm	CBC Flue Gas Sulfur lb/hr	Sp. G Ratio F/C	S Ratio F/C
11	0830	C		0.53				15.4		
	0947	F	3.62							11.
		C		0.32		1.175		17.6		
	1200	F	1.04		1.247				1.07	2.8
		C		0.36		1.166	1.1	6.3		
	1420	C		2.76				0.7		
	1515	F	3.46					0.3		~1.
		C		3.48						
	1630	F	3.32					0.3		
									$\overline{1.02}$ avg.	

APPENDIX K

FLY ASH ANALYSES, RUN 168-H

APPENDIX K

CBC FLY ASH ANALYSES, RUN 168-H *

DAY	8	9	10	11
HOUR	2030	1040	1730	1630
	Weight, %			
Constituent				
Fe (as Fe_2O_2)	15.8 (22.6)	13.5 (19.3)	14.0 (20.92)	22.3 (31.9)
Si (as SiO_2)	9.71 (20.74)	8.6 (18.37)	8.6 (18.37)	10.49 (22.41)
Al (as Al_2O_3)	6.27 (11.8)	5.41 (10.2)	5.9 (11.1)	12.5 (17.41)
Ca (as CaO)	16.22 (22.71)	19.63 (27.49)	22.1 (39.95)	24.38 (0.75)
S (as SO_3)	1.1 (2.75)	2.05 (5.12)	1.08 (2.7)	0.75 (1.87)
C	(12.9)	(18.5)	(13.1)	(6.94)
TOTAL	-- 93.5	-- 98.98	-- 96.24	-- 100.00

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*See page 6-26

APPENDIX L

COST ESTIMATE FOR RV-III DESIGN

Encl. 2

PROJECT Fluidized Coal Fired Boiler	**BY**	**DATE**	**CKD.** **DATE**
LOCATION RV III	**CONTRACT NO.**		
SUBJECT Cost Estimate	**SHEET NO.** L-2		**OF**

— STRUCTURAL SUPPORTS & PLATFORMS

	WEIGHT #	MTL COST $	FAB COST $	TOTAL COST $
1. MAIN BLR COLS. 6-10WF49 × 45'	13,230	1590.00	165.00	1755.00
" " BEAMS 2-24WF145 × 17	4,930	590.00	61.50	651.50
LWATIE BEAMS 6-16WF50 × 45'	13,500	1620.00	164.00	1784.00
COL. . 6-12WF36 × 17'	3,672	440.00	47.00	487.00
OTHER BLR MISC. TIES & etc APPR.	10,000	1010.00	[illegible]	1125.00
2. DUST COLL., AIR PRHTR, & OTHER MISC SUPP. APPR.	25,000	3000.00	312.50	3312.50
3. PLATFORM PLS APPR. 2000 # @ AVG 15#/#	30,000	3600.00	375.00	3975.00
TOTALS	100,332 50 TN	11,840.00	1,255.00	13,095.00

L-2

PROJECT Fluidized Coal Fired Boiler	BY RE DATE CKD. DATE
LOCATION RV III	CONTRACT NO.
SUBJECT Cost Estimate	SHEET NO. L-3 OF

MAIN BLR., CBC, & REG: PR. PARTS, INSUL., CASING & GRID PL. (1600 PSIG DESIGN)

	WEIGHT #	HTL COST $	FAB. COST $	TOTAL COST $
1. MAIN BLR. DRUM: 54" I.D. × 51'-3½" T-T (53'-10" O.A.)				
$t_r = \frac{2721600}{17500 \times 1.0 - [illegible]} = \frac{43200}{[illegible]} = 2.612''$ → SEMI-ELL'D				
[USE 2¾" SHELL (100% [illegible]) & 2¼" HDS.]				
SHELL: $\frac{56.75}{12} \pi \times 61.292 \times 112.2$ = 85,500				
HDS: 2 CIRC 65" DIA: 2 × [illegible].65 × 112.2 = 4400				
NOZZ., STUBS & SADDLES ---- 6,000	95,900	—	80,000	80,000
DRUM INTERNALS ------ 10,100	10,100	—	30,000	30,000
2. 2-(18" O.D. × 1.375) HDRS × 45'-3½" LG @ 244#/FT.	22,000	—	15,000	15,000
3. 3½" O.D. × .280 TUBES @ 9.65#/FT.				
SIDE WALLS: 2 × 73 × 15' = 2190'; 146 TUBES				
END do: 2 × 23 × 13' AVG = 598; 46 do				
INNER do: 3 × 71 × 14' do = 2982; 213 do				
TAKE-OFFS: 3 × 71 × 11' do = 2343; 213 do				
TOTALS 8113' 618 do	78,400	—	55,000	55,000
4. MEMBRANE FINS: ¼" × 2½": (2 × 72 × 11) + (2 × 24 × 13) + (3 × 72 × 7) = 3720' @ 2.5#/FT. INCL. WELDS	9,300	— $1/#+	11,000	11,000
5. 3-(8⅝" O.D. × .875) HDRS × 10.5' LG @ 72.42#/FT.	1,500	—	1,500	1,500
6. GRID PL. W/ [illegible] HOLES for SWAGED AIR NOZZ. (48 @ 2'-3" × 35'-8") + (81'-6" × 12') + (194' × 12') → 4460 ft²	9,200 30¢/#	10,000	$3/#+ 25,000	35,000
7. GRID AIR NOZZ. @ 100/FT., & @ 40¢/NOZZ.	1,400	18,000	$1/NOZZ. 45,000	63,000
8. CASING: (2 × 17' × 43'-3½") + 2([illegible]) → 1635 ft²	11,500	1,500	$3/# 5,000	6,500
9. INSULATION: 4" THK @ 6#/ft²; 1635 ft²	10,000	—	$3/# 5,000	5,000
10. CBC & REG TUBES (2½" O.D. × .200), @ 5#/FT; 5" P/R, 2 WALLS: 2(30 × 15' AVG) = 900'	4,500	—	5,000	5,000
11. SEC. SUPHTR. (12¾" O.D. × 1.000) INLET & OUTLET HDRS 2 @ 37' @ 125.5#/FT ------	9,300	—	5,000	5,000
ATTEMPERATOR [illegible] HDR: SAME AS ABOVE @ 30' LG + ATTEMP. SPRAY NOZZ., CONTROLS & ETC.	3,500	—	30,000	30,000
2" O.D. × .165 TUBES: 2¼Cr, 1 Moly; 5 ROWS @ 72/ROW (8% TO CARRY) @ AVG [illegible] → 5040' @ 3.2#/FT	16,300	8,000	8,000	16,000
12. MISC. DOWNCOMERS, FIT-UPS & ETC.	15,000	—	—	50,000
13. SUPER DUTY REFRACTORY (48' × 12' × 4½") = 216 FT³ @ 150#/FT³, @ $150/TON ---	32,400	2,500	330	2,830
TOTALS	330,900 165 TON			310,830

POPE, EVANS AND ROBBINS

PROJECT *Fluidized Coal Fired Boiler*	BY *RE*	DATE	CKD. DATE
LOCATION *KY III*	CONTRACT NO.		
SUBJECT *Costville*	SHEET NO. L-4L-3	OF	

PRIMARY SUPHTR. : (LOCATED ABOVE MAIN BLR) ; 1600 PSIG DESIGN

	WEIGHT #	MATL. COST $	FAB COST $	TOTAL COST $
1. INLETS: 21-2½" O.D. x .280 STUBS WELDED TO STEAM DRUM (21 to CARRY)				
2. ELEMENTS: 13 ROWS of 84-2½" O.D. x .280 BARE TUBES x 13' AVG. → 14,200' @ 5#/FT.	71,000	—	—	36,000
3. OUTLET HDR: 12¾" O.D. x 1.000 x 37' @ 125.5#/FT	5,000	—	—	3,000
4. ENCLOSURE: 2(6'x36')+2(6'x12') → 576 ☐' @ 14#/☐'	7,000	1,000	2,000	3,000
5. INSULATION: 3" THK @ 4½#/☐' → 576 ☐'	2,500	—	2,000	2,000
6. STRUCT. SUPPORTS: INCL ON SHT. C-1	—	—	—	—
TOTALS	85,500 43 TON			44,000

POPE, EVANS AND ROBBINS

PROJECT Fluidized Coal Fired Boiler	BY [illegible] DATE CKD. DATE
LOCATION RV-II	CONTRACT NO.
SUBJECT Cost Estimate	SHEET NO. C-5 OF

C-5

ECONOMIZER: 3 COMPONENTS: HO, REG & CBC (1600 PSIG DESIGN)

	WEIGHT #	MTL. COST #	FAB. COST #	TOTAL COST #
1. PORTION ABOVE "REG" FURNACE SECTION) 10 ROWS of 4-2½" S. #80 (.276 WALL) STD PIPE, SAIDS, BARE ELEMENTS (8 TO CARRY)				
2. INLET HDR: 1-8⅝" O.D. X .875 WALL x 3' LG. W/6" WATER INLET CONN. @ 72.42#/FT +	300	—	—	300.00
3. 2½" PIPE: 10x4x13' = 520' @ 7.66#/FT. —	4,000	—	—	2,000.00
4. OUTLET HDR & INLET HDR for CBC SECT. 1-8⅝" O.D. X .875 WALL x 8' LG — — — —	580	—	—	450.00
5. PORTION ABOVE "CBC" FURNACE SECT. 35 ROWS of 8-2½" PIPE (8 to CARRY): 35x8x13' = 3640' @ 7.66#/FT. — — —	2,800	—	—	1,400.00
6. OUTLET HDR. W/6" WATER OUTLET, 8⅝" O.D. x .875 WALL x 6' LG — — — — —	600	—	—	600.00
7. INLET HDR. W/6" WATER INLET 1-8⅝" O.D. x .875 WALL x 37-0" LG. @ 72.42#/FT — — — —	2,700	—	—	1,400.00
8. 22 ROWS of 84 2½" BARE PIPE (8 to CARRY) 22x84x13' = 24,024' @ 7.66#/FT	184,000	—	—	92,000.00
9. OUTLET HDR (SAME AS INLET HDR)	2,700	—	—	1,400.00
10. SUPPORTS (INCL. ON SHT C-1)	—	—	—	—
11. ENCLOSURE & CASINGS: APPROX 760# + 640# → 1400# @ 12#/# — — —	16,800	2,000	4,200	6,200.00
12. INSULATION 3" THK @ 4½#/ft → 1400 ft	6,300	—	4,200	4,200.00
TOTALS	220,780 ≈ 110 TON	—	—	109,950.00

POPE, EVANS AND ROBBINS

PROJECT Fluidized Coal Fired Boiler	**BY** CE **DATE** **CKD.** **DATE**
LOCATION RV II	**CONTRACT NO.**
SUBJECT Cost Estimate	**SHEET NO.** L-6 **OF**

TOP CASING, DUCTS, AIR PLENUM, & etc

	WEIGHT #	MAT'L COST $	FAB. COST $	TOTAL COST $
1. MAIN BOILER TOP CASING; (10'x12') + 2(AVG 5'x 36') → 480 ft² @ 7#/ft²	3,500 12¢/#	420 2/ft²	1,000	1,420
2. AIR PLENUM (Good for 48" wg.): 2(3'x43') + (12'x43') → 774 ft² @ 20#/ft²	15,500 12¢/#	1,850 3/ft²	2,300	4,150
3. MISC BREECHINGS, AIR DUCTS, & etc: APPROX. 4000 ft² @ 7#/ft²	28,000 12¢	3,360 2/ft²	8,000	11,360
TOTALS	48,000 24 T			16,930

POPE, EVANS AND ROBBINS

PROJECT Fl[illegible]ized Coal Fired Boiler	BY P.E. DATE	CKD.	DATE
LOCATION [illegible]	CONTRACT NO.		
SUBJECT Cost Estimate	SHEET NO. L-17		OF 2

— LIST of BOILER OPENINGS

1. Boiler Drum Access Opngs. (H.H. w/Inside, [illegible]) 2-12"x16"
2. Superheated Steam Outlet — — — — — 1-10"
3. Safety Valves: (On Suphtd Steam Outlet [illegible] 1-1½"
(On Boiler Steam Drum 1-1½"
do do do do 1-2½"
4. Level Bridle Conn. (On Boiler Steam Drum) — — — — 4-2"
5. Feedwater do (do) — — — — 1-6"
6. C.B.D. do (do) — — — — 1-1"
7. Vents do (do) — — 2-1"
8. Pr. Ga. do (do) 1-1"
9. Chem Feed do (do) — — — — 1-1"
10. Bottom Blow-Offs (On Lower Boiler Hdr.) — — — — 6-1½"
11. Drain Opngs (On Each Suphtr & Econ. Hdr.) — — — — 8-1"
12. Vent (do) — — — — 8-1"
13. Pr. Ga. (do) — — — — 4-¾" or ½"
14. Temp. Conn. (do) — — — — 8-1" or ¾"

Note:

All Opngs (2 thru 14) May Be
(a) All [illegible] Nozzles (400#, [illegible]) R.F. or R.J.
or (b) All Weld End Stubs.

* [illegible] The Cost [illegible] Is Preferred

PROJECT Fluidized Coal Fired Boiler	**BY** PC **DATE** **CKD.** **DATE**
LOCATION RV III	**CONTRACT NO.**
SUBJECT Cost Estimate	**SHEET NO.** L-8 **OF** [illegible]

BOILER TRIM (for 1600 PSIG DESIGN)

	#
1. @ SEC. SUPHTR. OUTLET: 1-10" ANGLE NON-RETURN (45°?)	7,000.00
1-10" SHUT-OFF VALVE W/ WARM-UP BYPASS	7,000.00
2. FEEDWATER @ ECONOMIZER INLET: 1-6" SHUT-OFF VALVE	2,500.00
1-6" STOP-CHECK VALVE	2,500.00
3-ELEMENT FEEDWATER REGULATOR: 1-6" VALVE 3000.00 POSITIONER [illegible]00.00 HANDWHEEL [illegible]00.00 INSTRUMENTS, TRANSMITTERS [illegible]	6,850.00
2-6" ISOLATING VALVES @ 2,500 EA —	5,000.00
1-2½" or 3" BYPASS VALVE — — —	1,000.00
3. 2-WATER COLUMNS W/ HIGH-LOW ALARMS, REMOTE TRANSMITTERS, ETC. @ 2000.00 EACH — — —	4,000.00
4. 6-1½" BLOW-OFF TANDEMS @ 200.00/VALVE, 400.00/TANDEM —	2,400.00
5. OTHER MISC. VALVES FOR: VENTS, SUPHTR & ECON. HDR DRAINS, CHEMICAL FEED, PR. GA. & ETC.:	
12-1" @ [illegible] EACH	1,200.00
24-½" [illegible] @ 50.00 EACH	1,200.00
6. POP SAFETY VALVES: (ALL CONSOLIDATED)	
1-1½" #1718D SET @ 1500 PSIG ON SEC. SUPHTR. OUTLET	1,843.00
1-1½ 1718A do @ 1600 do do BLR STEAM DRUM	1,427.00
1-2½ 1738A do @ 1600 do do do do do	1,783.00
	# 46,703.00

GUESSTIMATED WEIGHT [illegible] 3 TON

POPE, EVANS AND ROBBINS

PROJECT Fluidized Coal Fired Boiler	BY	DATE	CKD. DATE
LOCATION Ph III	CONTRACT NO.		
SUBJECT Cost Estimate	SHEET NO. L-9	OF	

F. D. FAN

CLARAGE FAN: (per DICK WESTON of WESTON ASSOC. INC. DET., MICH.)

D.I.D.W. - 54" WHEEL [illegible] VANES for CONTROL
APPROX. 15% OVERCAP. → 84,000 C.F.M. @ 70°F = [illegible] #/HR OF AIR
40.0 w.g. STATIC PRESS.
1775 R.P.M.
636 B.HP.

	COST $	WEIGHT #
FAN	7,200.00	8,000
DRIVE MOTOR: 700 HP, 1775 R.P.M., 2300V, 60~, 3Ø	14,000.00	4,000
TOTALS	21,200.00	12,000 6 TON

NOTE:
THE ESTIMATED MINIMAL REQUIREMENTS ARE:

330,000 #/HR of AIR = 73,400 C.F.M. @ 70°F.
31.0 w.g. STATIC PRESS.

WHEREUPON, THE ACTUAL BHP @ 1775 R.P.M. IS:

$$636 \frac{73,400}{84,000} \times \frac{31}{40} = 430 \text{ BHP}$$

PROJECT Fluidized Coal Fired Boiler	BY J.B. DATE CKD. DATE
LOCATION RV III.	CONTRACT NO.
SUBJECT Cost Estimate	SHEET NO. L-10 OF

L-10

	Equip.	Installed Cost
8. Dust Collector (Buell Eng. Co.)		
Main Boiler Reinjection	11,500	23,000
Regenerator Collector	1,400	2,000
OBC Collector	[illegible]	5,000
Precipitator	[illegible]	[illegible]
		$2[illegible],000
9. Bed Moving	2,000	4,000
10. Coal Supply		
Crusher	15,000	20,000
Mass flow conveyor - 11"	28,680	36,000
Redler, 50 ft. long, 10 H.P.		
[illegible]	[illegible]	[illegible]
48 Chute, [illegible] pipes, [illegible]	3,690	7,200
2 Air headers		400
		$65,700
11. Lightoff Burner	[illegible]	1,200
Regulator / Piping	1200	1,800
		$3,000
12. Ash Moving		
Main Collector - 8" Rotary, 5 H.P.		800
REG " - 4" " , 1 H.P.		[illegible]
OBC " - 8" " , 5 H.P.		800
Coal Makeup - 4" " , 1 H.P.		500
Piping and Air Supply		[illegible]400
		$[illegible],000

PROJECT	Fluidized Coal Fired Boiler	**BY** TB	**DATE**	**CKD.**	**DATE**
LOCATION	RY III	**CONTRACT NO.**			
SUBJECT	Cost Estimate	**SHEET NO.** L-11		**OF**	

13. Controls & Instrumentation

Combustion controls	$ 40,000
Feedwater / Attemperation	10,000
Instrumentation	20,000
	$ 70,000

14	Air Preheater, Ljungstrom 22 VIR-66 (Air Preheater Corp.)	109,000 equip. 30,000 inst. $ 139,000
15	Boiler Piping	20,000
16.	Boiler Electrical	20,000
17.	Building increment 30 ft. wide x 70 ft. long x 50 ft. high 105,000 ft³ @ $1.00	105,000

POPE, EVANS AND ROBBINS

APPENDIX M

LETTER FROM CE INDUSTRIAL BOILER OPERATIONS

INDUSTRIAL BOILER OPERATIONS. A DIVISION OF COMBUSTION ENG
WINDSOR, CONN. 06095

REPLY TO 1829 K STREET, N.W., WASHINGTON, D.C. 20006

M-2

CE INDUSTRIAL BOILER OPERATIONS

June 18, 1971

Mr. John W. Bishop
Pope, Evans & Robbins
515 Wythe Street
Alexandria, Virginia 22314

Subj: Monongahela Power
Rivesville, West Virginia
MAT 71406

Gentlemen:

This is in reply to your request for a budget price on the following equipment.

The unit that we are offering is arranged to burn pulverized coal and produce steam at 300,000 pounds per hour at 1270 psig and 925 degrees F with feedwater supply at 385 degrees F. Physical arrangement of the unit offered is shown on the attached marked-up drawing EP-683-04-1.

The equipment included in our offering is as follows:

- Boiler
- Superheater
- Economizer
- Structural Steel
- Gas Ducts
- Raymond Mills
- Ljungstrom Air Heater
- Feedwater Regulator
- C-E - APCS System
- Soot Blowers
- Service Representative
- Setting
- Erection Superintendence
- Waterwalls
- Desuperheater
- Casing & Buckstays
- Air Ducts
- "T" Burners
- Coal Piping
- Combustion Controls
- Steam Temperature Controls
- Burner Control Equipment
- Forced Draft & Induced Fans with Motor Drives
- Insulation & Lagging
- Freight to Rivesville,

The price for the above equipment delivered and erected.

Three Million Five Hundred and Thirty Eight Dollars----------$3,538,000.

The above price is a close budget estimate and of necessity the erection part of this price is based on average conditions.

Please let me know if we can be of any further service to you.

Sincerely,

Robert D. Taft

Robert D. Taft

RDT:db

APPENDIX N

WEIGHT AND COST ESTIMATE DATA

300 MW BOILER

PROJECT		**BY** RE **DATE**	**CKD.** **DATE** 6/18/71
LOCATION		**CONTRACT NO.**	
SUBJECT	300 MW BOILER	**SHEET NO.** N-2	**OF**

— WEIGHT & COST EST. DATA —

	WEIGHT #	MTL COST $	FAB COST $	TOTAL COST $
1. UPPER BLR. DRUMS: 2@60" I.D. × APP. 108' LG. SA-302B (M_n-½Mo) W/S = 20,000 @ UP TO 750°				
(a) SHELL: $\frac{6.5\pi}{12}$ × 108 × 204 × 2	750,000	—	$^{.60/\#}$ 464,000	
(b) HEADS: 4@6' DIA.: 28.274 × 204 × 4	23,000			
(c) EXTRA OPNGS: 2017 TUBE STUBS	—	—	^{45} 91,000	
(d) DRUM INTERNALS	60,000	—	180,000	735,000
2. LOWER BLR. DRUMS: 4@24" O.D. × 2.062 × APPROX 104' LG. @ 484#/FT. SA-106C W/S = 16,600 @ 700°F.				
(a) SHELLS: 104' × 484 × 4	201,000			
(b) HEADS: 8@2.75' DIA.: 8 × 5.94 × 84.2	4,000	—	$^{1.15}$ 235,000	235,000
(c) OPNGS., STUBS & ETC INCLUDED IN (a),(b)	—			
3. DOWNCOMERS: 8-18" O.D. × 1.562 × 20' ± LG. SA-106C W/S = 16,600 @ 700°F @ 275#/FT.				
(a) 8 × 20' × 275	44,000	—	$^{1.15}$ 50,000	50,000
4. BOILER TUBES: 3¼" O.D. × .340 (SA-209 T1a) (C-½Mo) @ 10.55#/FT; 16.5' AVG LGTH.				
(a) $(\overset{2017}{9\times193+5\times8\times7})$16.5 → 33,300 LIN. FT.	352,000	$^{2.89}$ 96,000	$^{30/\text{TUBE}}$ 61,000	157,000
5. MEMBRANE FINS: ¼ × 2¾ @ 2.34#/FT				
(a) [(9×192)8.625' + (5×8×8)16.5'] → 20,200 LIN. FT	48,000	5,000	$^{3/\text{FT}}$ 61,000	66,000
6. GRID PℓS W/DRILLED HOLES FOR SWAGED AIR NOZZ.				
(a) 2540 + 270 + 55.5 = 2,865.5 ft² of ½" Pℓ	60,000	$^{20/\text{ft}^2}$ 58,000	$^{50/\text{ft}^2}$ 144,000	202,000
7. GRID AIR NOZZ @ 100/ft²; 40¢/NOZZ; 32/#	9,000	115,000	$^{1/\text{NOZZ}}$ 287,000	402,000
8. CASING: (2×21'-9×96') + (2×21'-9×32') → 5570 ft²	39,000	$^{.12}$ 5,000	$^{3/\text{ft}^2}$ 17,000	17,500
9. INSULATION: 4" THK @ 6#/ft²; 5570 ft²	34,000	—	$^{3/\text{ft}^2}$ 17,000	17,000
10. SUPER DUTY REFRACTORY: 1000 ft³; 150#/FT.³; $150/TON	150,000	11,000	$^{40/\text{TON}}$ 3,000	14,000
TOTALS	1,774,000 887 TON			1,895,500

POPE, EVANS AND ROBBINS

PROJECT	BY RE DATE CKD. DATE 6/10/71
LOCATION	CONTRACT NO.
SUBJECT 300 MW BOILER	SHEET NO. - N-3 OF

N-3

— WEIGHT & COST EST. DATA —

	WEIGHT #	MTL COST $	FAB. COST $	TOTAL COST $
11. SEC SUPHTR: STEAM IN @ 700°F+, OUT @ 1000°F.				
(a) INLET HDR.: 24"O.D. x 2.343 x [illegible]'LG. @ 542#/FT. SA-209 T1A W/S = 14,400 @ 800°F	33,000	—	1.15 37,000	
(b) TUBES: 2.375 O.D.; (19 x 110 x 34') = 71,000 LIN. FT.				
I. 23,000' of SA-213 T11 x .250 WALL @ 5.673 #/FT.	130,000	2.15 50,000	—	
II. 24,000' of SA-213 T22 x .343 " @ 7.45 do	180,000	2.76 64,000	—	
III 24,000' of SA-213 TP316H x .343 @ 7.45 do	180,000	21.97 530,000	—	
IV 1980-180° U-BENDS @ 2#/EA & AVG $20/EA.	4,000	40,000	50,000	
(C) OUTLET HDR: 20"O.D. x 1.968 x [illegible]0'LG @ 379#/FT. SA-213 TP304 W/S = 13,750 @ 1000°F.	23,000	—	3.00 69,000	842,000
12. REHEATER: 700#/□ DESIGN; STEAM IN @ 650°F OUT @ 1000°F				
(a) INLET HDR.: 24"O.D. x .500 x 35'LG. @ 125#/FT. SA-106C W/S = 16,600 @ 700°F.	4,400	—	1.15 5,000	
(b) TUBES: 2.375 O.D.; (19 x 60 x 34') = 39,000 LIN. FT.				
I. 13,000' of SA-213 T11 x .154 WALL @ 3.652 #/FT.	48,000	2.15 28,000	—	
II 13,000' of SA-213 T22 x .154 " "	48,000	2.76 36,000	—	
III 13,000' of SA-213 TP316H x .154 " "	48,000	21.97 290,000	—	
IV 1080-180° U-BENDS @ 2#/EA; & AVG $20/EA.	2,200	22,000	28,000	
(C) OUTLET HDR.: 24"O.D. x .687 x 35'LG. @ 171#/FT. SA-213 TP304 W/S = 12,150 @ 1050°F.	6,000	—	3.00 18,000	427,000
13. PRI. SUPHTR.: STEAM IN @ 674°, OUT @ 760°				
(a) TUBES: 2.375 O.D. x .250; (11 x 170 x 34') = 64,000' @ 5.673 #/FT.; SA-213 T11 W/S = 13100 @ 900°F.	364,000	2.16 138,000	—	
(b) 1700-180° U-BENDS @ 2#/EA.; & $12/EA.	3,400	20,400	25,600	
(C) OUTLET HDR.: 24"O.D. x 2.343 x 90'LG @ 542#/FT. SA-209 T1A W/S = 14,400 @ 800°F.	49,000	—	1.15 56,000	240,000
14. ATTEMPERATOR: HDR. SAME AS (13C) EXCEPT APPROX. 20'LG. + SPRAY NOZZ'S, CONTROLS & etc.	11,000	40,000	1.15 13,000	53,000
TOTALS	1,134,000 567 TON			1,562,000

POPE, EVANS AND ROBBINS

PROJECT	BY RE DATE CKD. DATE 6/18/71
LOCATION	CONTRACT NO.
SUBJECT 300 MW BOILER	SHEET NO. N-4 OF

N-4

WEIGHT & COST EST. DATA

	WEIGHT #	MTL COST $	FAB COST $	TOTAL COST $
15. ECONOMIZER: FEEDWATER IN @ 480°F OUT @ 620° F.; MIN. DESIGN PRESS. - 3200 PSIG.				
(a) INLET & OUTLET HDRS.: 14" O.D. x 1.25 @ 171 #/FT. SA-106C W/S = 17,500 @ 650° (2x90') + (1x12') + (1x15) + (1x5') → 212'	36,000	←—— 1.15	43,000	
(b) TUBING: 2.875 O.D. x .276 @ 7.66 #/FT. (129,540 + 20,296 + 3400) = 153,236 LIN. FT.	1,180,000	2.15 330,000	——	
(c) 4200 - 180° U-BENDS @ 2 #/EA.; & $5/EA	8,400	21,000	60,000	
(d) ENCLOSURE & CASINGS 3100 ☐ @ 12 #/☐	34,000	.12 4,000	3/# 10,000	
(e) TOP CASING - 3300 ☐ @ 7 #/☐	23,000	.12 3,000	10,000	
(f) INSULATION: 3" THK @ 4.2 #/☐; 6400 ☐	29,000	——	3/# 19,000	471,000
16. MISC. BREECHING, AIR DUCTS & etc. APPROX 20,000 ☐ @ 7 #/☐	140,000	.12 17,000	3/# 60,000	77,000
17. AIR PLENUM (Good for 40" wg.): (a) 2(4'x100') + (32'x100') → 4000 ☐ @ 20 #/☐	80,000	.12 10,000	5/# 20,000	30,000
18. STRUCTURAL STEEL SUPPORTS: a) TOTAL at APPROX	1,000,000	——	.15/# 150,000	150,000
19. BOILER TRIM	50,000	200,000	——	200,000
20. F.D. FAN: MIN. REQMT'S: 2,204,400 #/HR = 490,000 cfm @ 70° 31" wg. STATIC PRESS. CLARAGE: SIZE 13161 DWDI; 80¾" WHEEL 1200 R.P.M. → 3000 BHP W/INLET VANE CONTROLS & etc 3000 HP/1200 R.P.M. DRIVER	32,000	36,000	——	36,000
TOTALS (LESS F.D. FAN DRIVER)	2,612,400 1306 TON			964,000

		Equipment	Installation Cost
21.	Dust Collectors (Buell Engineering Company)		
	Main Boiler Reinjection	$110,000	$150,000
	Regenerator Collector	2,400	5,000
	CBC Collector	11,000	20,000
	Precipitator	525,000	650,000
			$825,000
22.	Bed Moving	20,000	$40,000
23.	Coal Supply		
	Crusher	40,000	50,000
	2-120 ft Rollers -17'	116,000	150,000
	4-100 ft, 4" metering screws	10,000	20,000
	112 chutes, feed pipes, mushrooms	30,000	50,000
	4 air headers	1,200	4,000
			$274,000
24.	Lightoff Burners	8,000	$12,000
25.	Ash Moving		
	20 - 4" Rotaries	12,000	18,000
	1 - 8" Rotary	1,000	1,500
	1 - 16" Rotary	4,000	5,000
	Ash Reinjection Piping	2,000	4,000
	Branch Connections	3,000	6,000
	Air Supply and Miscellaneous	2,000	4,000
			$38,000
26.	Air Preheater, Ljungstrom 31-V1-86	595,000	$800,000
27.	Miscellaneous Direct Boiler Piping and Electrical		$200,000
28.	Building Increment		
	(only) 60 ft wide x 120 ft long x 60 ft high = 430,000 ft^3 @ $1/ft=		$430,000

APPENDIX O

PRORATING FOR CASE I - FLUIDIZED BED

APPENDIX O

TABLE O-1

Prorating for Case I - Fluidized Bed

	TVA Proc. B 1000 MW New	Prorate to:	Factor:	Scaled down to Case I "35.7" MW	
Sulfur related (p. 10)	2,275,000	35.7 MW	$\frac{253}{2260}$	253,000	
Gas Load related (p. 10)	3,190,000	1.48 MW	do not use	82,000	est. Scrubber
Process Subtotal	5,465,000			335,000	
General facilities	1,126,000	On Process Subtotal		69,000	
Total Direct	6,591,000	(TVA 6,621,000)		404,000	
Engineering Design	400,000			70,000	est.
Contractors Fees & Overhead	809,000	15% on Total Direct		60,000	
Contingency	440,000	20% on Total Direct		81,000	
Total Proj. Inv.	8,270,000			615,000	(1969)
Escalation @ 10%				62,000	
1969 to 1971				677,000	

Case III - Fluidized Bed

	TVA Proc. B 1000 MW New	Prorate to:	Factor:	Scaled down to Case III "243" MW	
Sulfur related		243 MW	$\frac{900}{2260}$	905,000	
Gas load related		10.1	do not use	350,000	prorate from Case I
Proc. Subtotal	5,465,000			1,255,000	
Gen. facilities		On Process Subtotal		258,000	
Total Direct	6,591,000	(TVA 6,621,000)		1,513,000	
Engineering Design				151,000	
Contr. Fees & Overhead		15% on Total Direct		227,000	
Contingency		10%		151,000	
Total	8,270,000			2,042,000	(1969)
Escalation @ 10%				204,000	
1969 to 1971				2,246,000	

Prorating for Case II - Pulverized Fuel Furnace -

Process A, new. Use "existing" for prorating slope

	TVA Proc. A 1000 MW New	Prorate to:	Factor:	Scaled down to Case II 35.7	
Total Direct	6,095,000	35.7	$\frac{540}{6316}$	520,000	
Engrg. Design	370,000			90,000	est.
Contr. Fees and Overhead	745,000	15% on Total Direct		78,000	
Contingency	410,000	20% on Total Direct		104,000	
Total	7,620,000			792,000	+ 10% esc.
				79,000	
				871,000	
Electrostatic ppt.	1,300,000*	n = .75,	$\frac{1}{\left(\frac{1000}{35.7}\right)^{0.75}}$		
Credit			or $\frac{1}{12.2}$	107,000	+ 10%
				11,000	
Net total	6,320,000				
		Escalation 10% 2 years			(1971)

*On plant cost not process equipment only Op. costs of plant required on investment before investment credit and then operating credit is subtracted.

Case IV - Pulverized Fuel Furnace

	TVA Proc. A 1000 MW New	Prorate to:	Factor	Scaled down to Case II 35.7	
Total Direct	6,095,000	Prorate to 243	$\frac{2,210}{6,316}$	2,130,000	
Eng. Design	370,000		$\frac{22}{49}$	166,000	
Contr. Fee & Overhead	745,000	15% on Total Direct		320,000	
Contingency	410,000	10% on Total Direct		213,000	
Total	7,620,000			2,829,000	+ 10%
					3,112,000
Electrost. ppt	1,300,000*	n = .75			
Credit	6,320,000		$\frac{1}{\left(\frac{1000}{243}\right)^{0.75}}$		
		$\frac{1}{2.88}$		452,000	+ 10%
					497,000
				2,377,000	
	Escalation 10% - 2 yrs				
				2,615,000	(1971)

*On plant cost not process equipment only Op. costs of plant required on investment before investment credit and then operating credit is subtracted.

TABLE O-2

LIMESTONE TREATING
WET-SCRUBBING REMOVAL
CAPITAL COST ESIMATES

	Case I	Case II	Case III	Case IV
Type Combustion	Fluidized Bed	Pulverized Fuel	Fluidized Bed	Pulverized Fuel
Limestone Addition Point	Scrubber	Furnace	Scrubber	Furnace
Boiler Plating				
Coal, Total, M lb/hr	26.9	26.9	182	182
MW				
Steam, lb/hr	300,000	300,000	1,900,000	1,900,000
SO_2 Total lb/hr	1890	1894	12890	12930
To Treater lb/hr	1703	1890	11650	12900
Gas to Treater lb/hr	15200	358,000	104,000	2,430,000
% SO_2 (Vol)	5.51	0.24	5.51	0.24
Estimated Capital Cost Based on 1969 TVA report plus 10% escalation for 1971 before credits	$677,000	$871,000	$2,246,000	$3,112,000
Estimated Electrostatic Precipitator credit	-	($118,000)	-	(497,000)
Net Capital Cost After Precipitator Credit	677,000	753,000	2,246,000	2,615,000

TABLE O-3
LIMESTONE INJECTION
WET SCRUBBING PROCESS

Operating Plant Capital Investment Capacity	CASE I - FLUIDIZED BED $677,000 100,000 lb/hr			CASE II - PULVERIZED FUEL $871,000 400,000 lb/hr			CASE III - FLUIDIZED BED $2,246,000 1,900,000 lb/hr			CASE IV - PULVERIZED FUEL $3,112,000 1,900,000 lb/hr		
	Units, Annual	$/Unit	Annual Cost $	Units, Annual	$/Unit	Annual Cost $	Units, Annual	$/Unit	Annual Cost $	Unit, Annual	$Unit	Annual Cost $
Raw Materials Limestone, 95% CaCO$_3$	13.3 M ton	2.05/ton	$ 27,300	12.5 M ton	2.05/ton	25,600	90.0 M ton	2.05/ton	186,500	85.0 M Ton	2.05/ton	174,500
Direct Costs Operating & Supervision	10,500 man-hr	5 man-hr	52,500	10,500 man-hr	5/man-hr	52,500	14,500 man/hr	5 man/hr	72,500	14,500 man/hr	5/man-hr	72,500
Water	32,500 M gal	0.10/M gal	3,300	49,600 M gal	0.10/M gal.	5,000	222,000 M/gal	0.10/M gal	22,200	335,000 M gal.	0.10/M gal.	33,500
Electricity (p.1)	617,000 KWH	.0006/KWH	3,700	2,380,000	.006/KWH	14,300	4,020,000	.006/KWH	24,200	16,100,000	.006/KWH	96,500
Maintenance-Labor & Material (APCO Guide - 5% of Capital)			33,900			43,600			112,300			156,000
Subtotal Directs			120,700			141,000			417,700			533,000
Indirect Costs Overhead (APCO Guide 150% of labor)			78,800			78,800			108,800			108,800
Capital Charges (APCO Guide 18% of Capital)			122,000			157,000			405,000			560,000
Subtotal Indirects			200,800			235,600			513,800			668,800
Total annual operating costs before credits			321,500			376,600			931,500			1,201,800
Precipitator operating & investment credit			-			30,600			-			133,000
Total annual operating costs after credit			321,500			346,000			931,500			1,068,800
75% Load factor 6500 hours operation												

APPENDIX P

FURTHER INFORMATION ON "RECOVERY PROCESSES"

APPENDIX P - Further Information on "Recovery Processes"

The probability of satisfactory commercial operability is an important criterion in the selection of processes to illustrate the potential of a complete system for producing power economically and meeting environmental standards. Conceptual studies and bench-scale tests are valuable indicators of direction which may be taken. However, numerous problems may arise which can be defined only by continuous pilot or prototype plant demonstrations. Such problems for example include unexpected deterioration or loss of treating agents, side reactions which diminish yield or produce materials which plug equipment, unpredicted corrosion or process hazards, or erratic performance. In addition the demonstration units should operate at the projected plant conditions in order to represent the equilibria and kinetics which are to hold in large scale operations.

One process which does not yet have the [illegible] background of demonstrated operability but is of interest is the Magnesium Oxide process of the Chemical Construction Corporation. The particularly noteworthy feature is that the sulfur dioxide is captured as magnesium sulfite which is proposed to be regenerated to recover SO_2 and MgO for reuse. A prototype commercial system will be built at the Mystic No. 6 Unit of Boston Edison Co. to handle flue gases from a 155 mw boiler (5). The anhydrous spent crystals will be shipped to a separate location for processing to 98% H_2SO_4. Such an arrangement should be acceptable to power plants for several reasons: the chemical processing at the boiler house is minimized; the product being handled is not hazardous or obnoxious; and the need for disposal sites for spent agents is avoided. In the broad sense, public relations from a conservation standpoint should be good because the treating agent is recycled instead of being spent once-through and the SO_2 is conserved for further use.

The SO_2 capture part of the process appears to be very close to the limestone addition plus wet-scrubbing process, and pending the prototype plant results, it is reasonable to assume that investments and operating costs will be nearly the same. Some differences are foreseen in that the slurry will start with oxide rather than with carbonate. It is expected that the MgO process will require solid-free gas in order to preserve the purity of the agent. Provision of this cleaning will offset the precipitator credit considered for the conventional boiler with the limestone process where the scrubber can provide effective solids removal. The solids removal operation with the fluidized-bed boiler should be a minor matter due to the low gas volume, and its low solids content, supplied to the SO_2 removal system.

The benefits to the boiler operations may be limited to the intangibles described above. However, assuming that the spent magnesia is at least sufficiently [illegible] as a sulfur source that it will be exchanged at no charge for regenerated magnesia f.o.b., the reprocessing plant, the shipping cost for the exchange, should be about the same as the cost of purchased limestone.

For the 300,000 lb/hr fluidized-bed boiler the use of 13300 tons per year of limestone is anticipated. At $2.05 per tone delivered the limestone would cost $27,300 annually. Using magnesia and allowing 110% of the stoichiometric quantity, 4500 tons of regenerated magnesia would be broughtin and 10,000 tons of spent magnesia would be returned for SO_2 recovery and regeneration of the magnesia. The economics will require development following the prototype operation and firm estimates of the capture and recovery portions of the process. However, it appears that the total material to be shipped with the magnesia process is about the same as the limestone used in that process.

Some comparable background is available on the [illegible] portion of the magnesia process. Iowa State University (6) and Kent Feeds, Inc. studied the production of sulfuric acid from calcium sulfate by heating to produce SO_2, but definitive conclusions will require data on magnesia operations and their analysis under present conditions. It should be noted, however, that the recovery operations may be carried out as a separate chemical business which could accept spent magnesia from a number of boiler plants and carry out the marketing of the acid produced.

A process which has the background of commercial operation and developmental studies is direct reduction of SO_2 to sulfur. Allied Chemical Corporation (P-12) has reexamined the hot coke reduction process used at one time at Trail, B.C. by Consolidated Mining and Smelting Company (P-13). The study included 4 cases with coke as the reductant [illegible] SO_2 in the gas and treating [illegible] tons per day [illegible]. Allied Chemical estimated a fixed capital requirement of $5,200,000 and $42.68 net operating cost per ton of sulfur (330 days per year). The process would be of particular interest in connection with the fluidized-bed boiler since the low oxygen content of the treater gas minimizes coke requirement and the boiler could supply hot cokes at fuel value. In contrast, Allied Chemical based operating costs on outside coke at $24 per ton and found coke cost to be 32% of the total operating costs. Use of coke from the boiler would reduce operating costs by about 20%. On the basis of the Allied Chemical estimates, it appears that sulfur credits would not justify [illegible]tion with the present technology. However, there is an opportunity to consider possible economies in design if full advantage be sought in designing the reduction unit as an integral part of the fluidized-bed boiler.

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TABLE P-1
SYSTEMS STUDIED

Case	Size lbs/hr steam	Type of Combustion	Stream Treated
I	300,000	Fluidized-Bed	Regenerator Gas
II	300,000	Pulverized Fuel	Flue Gas
III	1,900,000	Fluidized-Bed	Regenerator Gas
IV	1,900,000	Pulverized Fuel	Flue Gas

TABLE P-2
FEED TO TREATING UNIT

Case	I	II	III	IV
Gas Flow lb/hr	15,200	358,000	104,000	2,430,000
SO_2 Flow, lb/hr	1703	1,890	11,650	12,900
Temperature, °F	450	350	450	310
Gas Analysis	(% Vol)			
O_2	1.02	3.94	1.02	3.94
SO_2	5.51	0.24	5.51	0.24
CO_2 %	16.78	13.49	16.78	13.49
N_2 %	68.29	74.64	68.29	74.64
H_2O%	7,40	7.69	8.40	7.69

GENERAL ASSUMPTIONS USED BY APCO FOR COSTING

1. Plant size: 1000 MW*

2. Load factor (two cases): 55% for existing plants
   75% for new plants

3. Percentage sulfur in coal: 3.5%

4. Fixed charges:
   7% depreciation
   3% taxes & insurance
   8% cost of money
   Total 18% (annual percentage of capital investment)

5. Variable charges:
   labor @ $5.00/hr + 150% overhead
   maintenance @ 5% annually of capital
   electricity @ 6 mils/kwh.
   fuel gas or oil @ 45¢/10^6 BTU
   coal @ [illegible]/10^6 BTU
   limestone @ $2.00/T delivered
   cooling water @ 10¢/1000 gal

6. Credits for by-products:
   acid (100%) @ $10/T
   sulfur @ $20/T

7. Heating value of coal: 11,800 BTU/lb

Power station efficiency: 34.1%, equivalent [illegible]

* Not applicable in this study

A. Ammonia Scrubbing (phosphate fertilizer): TVA

Status Design & cost study

Principle Ammonium sulfate produced is used with phosphate ore and nitric acid to produce fertilizer

Economic Aspect If pollution abatement is mandatory, there is economic promise, in fertilizer producing areas, for the larger boiler plants.

Reference Sources Slack (P-10)

Summary Possible utility with favorable combination of circumstances.

Cost (P-10)

B. Cat-Ox: Monsanto

Status Pilot, Prototype

Principle Catalytic conversion to sulfuric acid in gas.

Economic Aspect Requires disposal of [illegible] [illegible] acid. Predicted high cost of installation but low labor and utilities.

Reference Sources Chilton (P-1) Cortelyou(P-6) Slack (P-10) Stanford Research Institute (P-20)

Summary Handling and marketing problems of dilute acid limit acceptability

Cost (P-6, oil fuel)

C. **B&W - Esso Adsorption:** Babcock & Wilcox
Esso Research & Engineering

Status Pilot Prototype planned as cooperative effort with utility companies.

Principle SO_2 adsorbed on proprietary solid

Reference Sources [illegible]
Environmental Science and Technology [illegible]

Summary Details not available.

D. **Cominco Absorption:** Consolidated Mining and Smelting Company
Canada, Ltd.
Licensor, Olin-Mathieson Chemical Corporation

Status Commercial Type Combination

Principle: [illegible] ammonia. The resulting solution is acidified and stripped to produce a partly concentrated gas (about 24% SO_2). The stripped liquor contains ammonium sulfate which is crystallized as a by-product.

Economic Aspect: The enriched gas is suitable for further processing [illegible] to acid

Reference Sources Stack [illegible]

Summary Of greatest interest as part of a complex for acid and fertilizer operations.

Cost McKee (P-11)

E. Contact [illegible]

Status: Commercial (Smelters)

Principles: Hot gas is scrubbed [illegible] [illegible] and humidify it, then cooled, and dried. It is then passed through the converter and absorber system to produce 66 Be acid.

Economic Aspect: While a contact plant can be designed for low concentration of SO_2, three percent has been given (P-11) as a guide to the minimum for economic application to smelter gases. This concentration could be met with the rich (about 5.5%) SO_2 expected in the greater gas stream with the fluidized-bed boiler, after mixing with the required air for conversion. With the low concentration of SO_2 (about 0.24%) in conventional flue gas, the gas volumes to be processed could be much greater per unit of acid made.

Reference Source: McKee (P-11)

Summary: [illegible] large installation, [illegible] SO_2 and pollution abatement credit, economics possibly acceptable. Introduces acid technology and marketing problems to utility operations.

Cost: (P-11)

F. Potassium Polyphosphate Scrubbing: TVA

Status Under development **Type** Combination: Absorption and conversion to sulfur

Principle Scrubbing obtains potassium pyrosulfite $K_2S_2O_5$ (precipitates). From the $K_2S_2O_5$ a third of the sulfur is released by heat as SO_2 and the balance is obtained as H_2S by reduction. The H_2S and SO_2 are catalytically reacted to produce sulfur.

Reference Sources Slack (P-10)

Summary Process details and economics are available.

G. Princeton Chemical Research

Status Pilot plant

Principle A Claus-type reduction of SO_2 to elemental sulfur by hydrogen sulfide, generated by conversion of part of the sulfur with natural gas.

Summary Princeton Chemical Research has prepared investment and operating costs and have suggested further pilot studies and development.

Reference Sources Princeton (P-14)

H. Potassium Carbonate - Thiocyanate: Garrett Research & Development Co., Inc.
LaVerne, California

Status Laboratory

Principle A slurry of potassium carbonate in molten potassium thiocyanate (M.P. 172.3°C) is used to absorb SO_2 at 180°C. The spent agent is filtered and cake containing potassium sulfite blended with coal and roasted to produce K_2S. The melt is leached with water, and ash removed. The potassium carbonate is regenerated by carbonation, evaporation and calcining.

Economic Aspect Garrett estimates capital costs of $17,600,000 and net treating costs at $147 per ton of fuel for a 1300 mw power plant burning 3.5% S coal.

Reference Source Chemical and Engineering News, April 12, 1971, page 65

Summary Further information needed for evaluation

I. Nitrosyl Sulfuric Acid: Tyco

Status Bench Type Conversion to sulfuric acid

Principle Add nitrogen oxides to gas and absorb nitrosyl sulfuric acid in sulfuric acid following chamber acid technology

Reference Sources Chilton (P-1) Slack (P-18)

Summary Conceptual

L. **Formate:** Consolidation Coal Co., Library, Pa

Status Study **Type** Combination (Absorption plus reduction to H_2S for conversion to S)

Principle SO_2 is absorbed in concentrated solution of [illegible] further [illegible] and [illegible] with carbon dioxide and [illegible] being released as hydrogen [illegible]

Reference Sources Chilton (P-1), Environmental Science and Technology (P-[illegible])

Summary Details not available

M. Ionics/Stone & Webster: Stone & Webster Corp.
Ionics, Inc.

Status Pilot Type Combination (absorption plus electrolytic)

Principle Flue gas after precipitator is quenched (solids removed) and SO_2 absorbed in caustic solution. SO_2 is sprung by acidification and the liquor then electrolyzed to regenerate caustic and acid. The SO_2 product is processed to sulfuric acid.

Economic Aspect The process is acknowledged to use large amounts of power but the proponents (P-2) point out that provision of large surge tanks will permit regeneration with the electrolytic cells at off-peak hours. Some cost factors are given (P-2)

Reference Sources Humphries, et al (P-2)

Summary Applicability most likely on large central stations where 500 ton/day or more of acid can be produced. Wide range of technology involved.

J. Magnesium Oxide Scrubbing: Chemical Construction Corp.
Basic, Inc.

Status Prototype Type Absorption and recover of SO_2

Principle Absorption of SO_2 in slurry of MgO, generally similar to limestone process. Spent magnesia is separated, dried, and calcined to regenerate magnesia and recover SO_2. The process of going through a dry stage facilitates the installation of regeneration step at a separate location.

Economic Aspect Absorption expected to be similar to limestone process. Potential for recovery of SO_2 for further use.

Reference Sources Shah P-3) Tucker and Burleigh (P-4)

Summary Opportunity for sulfur recovery with minimum chemical processing at boiler plant.

K. DMA Absorption: American Smelting and Refining Company

Status Commercial Type Concentrates SO_2

Principles Utilizes dimethylaniline as an absorbent to produce highly concentrated (90% plus) SO_2

Economic Aspects Effectively concentrates SO_2 for other processing or purification for sale.

Reference Sources McKee (P-11) Allied Chemical (P-12)

Summary Demonstrated system, requires supplementary operations to process SO_2 to salable product.

Cost (P-12)

N. Limestone Addition, Wet Scrubbing: Various

Status Continuing study Type Throw away

Principle Ground limestone added to recirculating slurry stream at flue gas scrubbing system to react with SO_2 absorbed. Spent agent is discarded.

Economic Aspect No recovery of sulfur values but applicable for relief of SO_2 pollution at relatively low investment.

Reference Sources TVA (P-9)

Summary Principle was used commercially in England. Throw away process, disposal site required.

Cost (P-9)

O. Reinluft Activated Char: Dr. F. Johswich

Status Has been commercial Type Absorption and regeneration

Principle SO_2 absorbed on a descending bed of char. The char is regenerated thermally by an inert gas stream yielding SO_2 (approximately 50%) for use in an acid plant.

Economic Aspect High carbon consumption (P-10)

Reference Sources: Wiedersum (P-7), Cortelyou (P-6)
Slack (P-10), McKee (P-11),
Frankenberg (P-18)

Summary Has been reported (McKee (P-11) to be high cost and to have process hazards.

Cost (P-6) oil fuel

Wiedersum cites Furkert, H. Proc. Am. Power Conf. 32 (1970)

P. Limestone Injection, Wet Scrubbing: Combustion Engineering

Status Prototype, continuing Type Throw away

Principle Ground limestone is injected into combustion gas in boiler, being calcined and reacting with SO_2 during passage. Flue gas is then scrubbed with recirculated slurry of water plus solids removal from gas and the spent agent is discarded.

Economic Aspect No recovery of sulfur values but process applicable for relief of SO_2 pollution.

Reference Sources Chilton (P-1), Cortelyou (P-6) TVA(P-9)

Summary Throw away process but relatively simple to employ. Disposal site required.

Cost (P-6, oil fuel) (P-9)

Q. Limestone - Dry TVA
(or dolomite)
(or nahcolite (sodium carbonate))

Status Further testing (P-1) Type Throw away

Principle Alkaline acceptor injected into pulverized coal furnace. Spent agent discarded with fly ash.

Economic Aspect Low investment and relatively little effect on boiler economy or chimney plume.

Reference Sources Chilton (P-1) TVA (P-8)

Summary Low removal efficiency (20-70%) but possible utility as standby palliative for small units or where air pollution is marginal.

Cost (P-8)

R. Citrate Absorption: US Bureau of Mines, Salt Lake City, Utah

Status Pilot plant

Principle Gas is cooled to 50 °C, cleaned of solids and sulfuric acid. It is then scrubbed with sodium citrate solution which absorbs the SO_2 as a bisulfite-citrate complex. The spent solution is regenerated by hydrogen sulfide which precipitates sulfur. The sulfur-liquor slurry is centrifuged and concentrated slurry is heated at 130 °C under pressure to melt and settle out the sulfur. A portion of the sulfur is reduced to hydrogen sulfide for the process with natural gas and steam over an alumina catalyst.

Economic Aspect Bureau of Mines estimates $13,000,000 capital cost and $35 operating cost per ton of sulfur for smelter plant recovering 114,000 tons per year (95%) from 2% SO_2 gas. For conventional furnace stack gas at 0.24% SO_2 the gas volume/sulfur ratio would markedly raise costs. With 5.5% SO_2 as in the fluidized-bed boiler, there would be a better chance that sulfur credit would bring the cost in line with limestone processes for abatement.

Reference Sources Chemical & Engineering News (P-17)

Summary Process is interesting, further information necessary for comparison

Cost (P-17)

S. Reduction by Coke Consolidated Mining and Smelting Company of Canada, Lt.

Status Commercial background. Process also employed by Imperial Chemical Industries and Boliden

Principle The gas is treated with incandescent coke to reduce SO_2 to elemental sulfur. Supplementary processing is required to convert other sulfur compounds formed as by-products.

Economic Aspect The cost of purchased and shipped coke has penalized the process compared to similar processes employing natural gas or other gaseous reductants. Low SO_2 content is a negative factor as is high oxygen which consumes reductant.

Reference Sources Allied Chemical (P-12)

Summary The conditions existing with the fluid-bed boiler may provide a favorable situation. The SO_2 content is relatively high (5.5%), and oxygen is low (1%). In addition there is the potential for integrating the boiler with the treating system to utilize boiler coke and to return spent gas to the boiler to reduce the clean-up operations conventially required on the spent gas.

Cost Allied Chemical (P-12)

T. Alkalized Alumma Bureau of Mines

Status Suspended Type Combination (Dry adsorption and regeneration to H_2S)

Principle: Adsorption of SO_2 at about 600°F on basic sodium aluminates, regenerated by producer gas to form H_2S.

Economic Aspect Appears high cost due to attrition losses of adsorbent and magnitude of equipment for handling gas volumes.

Reference Sources Chilton (P-1) Cortelyou (P-6)

Summary Status uncertain

Cost (P-6, oil fuel)

U. Reduction by Natural Gas: Asarco

Status Semi-commercial background

Principle The natural gas (methane) is mixed with the SO_2 bearing gas and reacts with the SO_2 and oxygen in a combustion chamber to form elemental sulfur. Side reactions produce hydrogen sulfide and some carbonyl sulfide. The process requires extensive heat exchange and provisions for conversion of the by-products for sulfur recovery and cleanup of the vent gas.

Economic Aspect The process is effective. Positive factors for economy are high concentration of SO_2 and low concentration of oxygen which consumes reductant. The use of methane is reported as lower cost than purchased coke, and broadly comparable results are reported for other gaseous reductants such as reformed methane. For very large (200 tons or more per day of sulfur) and very favorable concentrations (15% SO_2, 1.2% O_2), Allied Chemicals reported figures show operations cost less than the $20 per ton value being used.

Reference Sources Allied Chemical (P-12) McKee (P-11)

Summary Allied Chemical has compared a number of reductants. The figures given indicate doubtful applicability of the process for conventional boilers although it could be worthwhile with more favorable systems.

Cost (P-11) (P-12)

V. Water (Alkalized) Washing: British

Status Dormant Type Throwaway

Principle Flue gas scrubbed with river water (alkalized with chalk) which is discarded.

Economic Aspect Effective but numerous problems with operation, waste disposal, loss of gas buoyancy.

Reference Sources Chilton (P-1) TVA (P-[illegible])

Summary Precursor of limestone plus wet-scrubbing systems.

W. Molten Carbonates Atomics International

Status Small-scale tests Type Combination (capture and conversion to S)

Principle Molten eutectic of lithium, sodium and potassium carbonates captures SO_x. Reducing gas converts sulfites and sulfates to sulfides. Steam is used to separate H_2S for Claus process conversion to elemental sulfur. High temperature (800°F or above) (P-7).

Economic Aspect Process and equipment development necessary.

Reference Sources Chilton (P-1), Wiedersum (P-[illegible]) Singmaster & Breyer (P-19).

Summary A pilot plant has been recommended to define problem areas and permit complete technical and economic evaluation.

X. Wellman-Lord Wellman-Lord Inc.

Status Prototype
(commercial on acid plant (P-5)

Principle SO_2 is absorbed in sodium sulfite solution to form bisulfite which is separated and decomposed to produce pure SO_2. (An earlier version used potassium sulfite to form pyrosulfite.)

Economic Aspect For conventional power plant of 500 mw size, the SRI figures indicate an operating cost of $80 per ton of sulfur recovered. For a smelter case comparable to Case III of the present study, the indicated cost is $55 per ton. It is doubtful that present sulfur credits will justify the selection of the process as presently designed.

Reference Sources Chilton (P-1), Env. Sci. & Tech (P-5) Stanford Research Institute (P-20).

Summary Further demonstration and evaluation needed to justify selection for abatement.

Y. **DAP-Mn** Mitsubishi Heavy Industries, Ltd.

Status Prototype **Type**: Combination (Capture and recovery as ammonium sulfate)

Principle Activated manganese oxide powder is injected into the flue gas and passed through a fluidized reaction chamber. It is stated that the oxide has a great affinity for SO_2 at 100°C to 180°C. Following the reaction chamber, the solids are collected and recycled, with a portion being withdrawn for regeneration. The regeneration consists of treatment with aqueous ammonia and air under pressure. The regenerated manganese oxide is filtered out for reuse and the ammonium sulfate solution sent to a crystallizer. Soot from the flue gas on the agent is said to be readily removable by flotation from the solution.

Economic Aspect Limited market for ammonium sulfate.

Reference Sources Chilton (P-1) Uno (P-24)

Summary While the production of ammonium sulfate limits acceptability, it may be noted that 90% removal of SO_2 is claimed from flue gas containing the low concentration of 0.11 vol% of SO_2.

APPENDIX Q

ARSENIC ANALYSIS DATA, RUN 171-H

APPENDIX Q

TABLE Q-1
ARSENIC ANALYSIS DATA, RUN 171-H
(As CONCENTRATIONS IN μg/g)

		Bed Material			REG	REG Bed
Day	Time	FBM*	CBC	REG	Fly Ash	%S
12	2035	5.5⁻				
12	2245			5.5		0.6
13	1045	3.9⁻		4.5		3.3
14	2250	6.1⁻		11.9		1.4
15	0745			11.6		1.7
15	0845	6.6⁻				
15	1245			7.2		0.24
16	1000	9.3⊕				
16	1100			10.3		0.6
16	2215	10.9⊕				
17	0900			16.8**		1.32
17	1130			17.2**		0.31
17	1535	8.9⊕	9.4			
17	1815	2.2***				
17	2045			0.***	6.8	0.35
18	1145			7.6	10.7	0.22
18	1815		9.5			
18	2100	8.4⊕				
18	2200				20.4	
18	2300	13.⊕		13.	13.5	0.56
19	1100	19.7⊕		7.5	10.	0.36
19	1300	6.5⁻		6.8	10.9	0.83
19	1510				22.8	
average		8.4	9.4	9.2	13.6	
median		8.3	9.4	7.6	10.9	

Notations - = below median, ⊕ = above median
** causes bias of average
*** after adding salt to FBM bed
Blank spaces indicate no analysis

APPENDIX R

CONVERSION FACTORS

TABLE R-1
CONVERSION FACTORS

The below-cited conversion factors are provided to assist readers who are more familiar with metric units than with the units used in this report. Although it is EPA's policy to use the metric system in all its documents, particularly those of a technical nature, this report reflects certain non-metric units utilized during the 1970-1971 experiments upon which it is based.

The non-metric unit:	Multiplied by:	Yields the metric:
Btu	252.00	cal
°F	5/9(°F-32)	°C
ft	0.30	m
ft^2	0.09	m^2
ft^3	28.32	l
gal.	3.79	l
gr	0.65	g
HP	746	W
in.	2.54	cm
$in.^2$	6.45	cm^2
lb	0.45	kg
ton	907.18	kg